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PROCESS DESIGN MANUAL FOR PHOSPHORUS REMOVAL

Black and Veatch

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27 **Abstract**

The discharge of phosphorus-containing wastewaters into the surface waters of the United States has contributed to their over-fertilization and eutrophication. As a result, efforts are now being made to remove phosphorus from wastewater.

This manual discusses phosphorus removal methods that have been found effective and practical for use at treatment plants. All the methods included involve chemical precipitation of the phosphorus and removal of the resultant precipitate. Precipitants include salts of aluminum and iron, and lime. The practical points of addition are before the primary settler, in the aerator of an activated sludge plant, before the final settler, or in a tertiary process.

Included in the discussion of each treatment method is a description of the method, pilot or full-scale performance data, equipment requirements, design parameters, and costs. This information should be of value to designers, municipal officials, regulatory agencies, city planners, and treatment plant operators.

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PROCESS DESIGN MANUAL FOR PHOSPHORUS REMOVAL

For

U. S. ENVIRONMENTAL PROTECTION AGENCY
Technology Transfer

By

BLACK & VEATCH, CONSULTING ENGINEERS
P. O. Box 8405
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Program #17010 GNP
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The mention of trade names or commercial products in this manual is for illustration purposes, and does not constitute endorsement or recommendation for use by the Environmental Protection Agency.

ABSTRACT

The discharge of phosphorus-containing wastewaters into the surface waters of the United States has contributed to their over-fertilization and eutrophication. As a result, efforts are now being made to remove phosphorus from wastewater.

This manual discusses phosphorus removal methods that have been found effective and practical for use at treatment plants. All the methods included involve chemical precipitation of the phosphorus and removal of the resultant precipitate. Precipitants include salts of aluminum and iron, and lime. The practical points of addition are before the primary settler, in the aerator of an activated sludge plant, before the final settler, or in a tertiary process.

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FOREWORD

The formation of the Environmental Protection Agency marks a new era of environmental awareness in America. This Agency's goals are national in scope and encompass broad responsibility in the area of air and water pollution, solid wastes, pesticides, and radiation. A vital part of EPA's national water pollution control effort is the constant development and dissemination of new technology for wastewater treatment.

It is now clear that only the most effective design and operation of wastewater treatment facilities, using the latest available techniques, will be adequate to meet the future water quality objectives and to ensure continued protection of the Nation's waters. It is essential that this new technology be incorporated into the contemporary design of waste treatment facilities to achieve maximum benefit of our pollution control expenditures.

The purpose of this manual is to provide the engineering community and related industry a new source of information to be used in the planning, design, and operation of present and future municipal wastewater treatment facilities. It is recognized that there are a number of design manuals, manuals of standard practice, and design guidelines currently available in the field that adequately describe and interpret current engineering practices as related to traditional plant design. It is the intent of this manual to supplement this existing body of knowledge by describing new treatment methods, and by discussing the application of new techniques for more effectively removing a broad spectrum of contaminants from wastewater.

Much of the information presented is based on the evaluation and operation of pilot, demonstration and full-scale plants. The design criteria thus generated represent typical values. These values should be used as a guide and should be tempered with sound engineering judgment based on a complete analysis of the specific application.

This manual is one of four now available through the sponsorship of the Environmental Protection Agency to describe recent technological advances and new information in the following subject areas:

Carbon Adsorption
Phosphorus Removal
Upgrading Existing Plants
Suspended Solids Removal

These manuals are the first edition copies and will be updated as warranted by the advancing state of the art to include new data as it becomes available, and to refine design criteria as additional full-scale operational information is generated.

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Chapter 1
INTRODUCTION

1.1 Purpose

The technology for removal of phosphorus from wastewater has developed rapidly in the last few years. The need for practical phosphorus removal procedures is a result of the over-fertilization and eutrophication of the Country's surface waters. This further resulted in the establishment of state water quality standards that limit the concentration of phosphorus in receiving waters. As an example, states that have streams tributary to the Great Lakes have established wastewater effluent phosphorus limits or require a fixed percentage reduction of phosphorus. Several other states have also established phosphorus standards.

The Environmental Protection Agency has sponsored many research and demonstration studies at several cities in the past few years to advance the knowledge of phosphorus removal. Local and state governments and private industries have also contributed to this work. This manual is intended to summarize process design information for the best developed removal methods that have resulted from this governmental and private effort.

1.2 Scope

This manual discusses a number of phosphorus removal methods that involve chemical precipitation. Phosphorus removal obtainable by biological activity alone is not included. Treatment methods in which phosphorus removal occurs, but is not a principal objective, are also omitted. The latter group of processes includes ion exchange, reverse osmosis and other demineralization treatments which at present are more closely associated with wastewater renovation and reuse than with pollution control. These will be included in updated versions of the manual when appropriate. One application of a precipitation method that has been omitted and which is of growing interest is complete lime treatment of raw sewage. This application will be included when the manual is updated.

The information included on chemical methods of phosphorus removal was obtained from the available literature, progress reports of demonstration studies, and private communications with investigators actively working in the field. Design guidelines have been developed from these sources.

The information contained in this manual is oriented toward design methods and operating procedures for removal of phosphorus from wastewater to comply with state water quality standards. It is recognized that present state standards may eventually be superseded with more stringent requirements for phosphorus removal. Such revisions in state standards must be considered in design of treatment facilities to minimize future modifications to treatment plant structures.

Engineers are often faced with circumstances that require the design of treatment facilities with a limited amount of data, testing, and time. Such circumstances require a more conservative design to allow for contingencies. Since most phosphorus removal methods are of recent development, they fall into this category. Conservative values are given, therefore, when stating design parameters.

Chapter 2

BASIC DESIGN CONSIDERATIONS

2.1 Sources of Phosphorus

Domestic wastewater normally has a substantial concentration of phosphorus with the primary sources being a result of man's activities in the home. Human wastes such as feces, urine, and waste food disposal account for about 30 to 50% of the phosphorus in domestic wastewater (1). Detergents containing phosphate builders and used principally for laundering of clothes account for the remainder of phosphorus, or about 50 to 70%. Other sources of phosphorus may cause deviation from these percentages. For example, where sodium hexametaphosphate or other phosphorus compounds are used as corrosion and scale control chemicals in water supplies, the phosphorus added will be present in the same concentration, although not necessarily in the same form, as in the water supply. This source can account for 2 to 20% of the total phosphorus present in the wastewater.

Industrial wastewater discharges to the sanitary sewer system may either dilute or increase the concentration of phosphorus present in the combined wastewater. For example, wastewaters from pulp and paper mills may be phosphorus deficient and therefore dilution will occur. On the other hand, wastewaters discharged from some industries, such as potato processing plants, may contain high concentrations of phosphorus and consequently may increase the phosphorus concentration of the combined wastewater.

The quantity of phosphorus resulting from human excretions reportedly ranges from 0.5 to 2.3 lb per capita per year (2). The mean annual excretion is estimated to be about 1.2 lb per capita. The mean annual contribution of phosphorus from synthetic detergents with phosphate builders is estimated to be about 2.3 lb per capita, at present. Thus, exclusive of industrial wastes and other phosphorus sources, such as water softening or sequestering agents, the domestic phosphorus contribution to wastewater is about 3.5 lb per capita per year.

On this basis, the wastewater phosphorus concentration may be estimated for a community where there are no phosphorus data available. Assuming a population of 5,000, the total phosphorus discharged equals 5,000 x 3.5 = 17,500 lb/year, or 48 lb/day. If a wastewater flow of 100 gal. per capita per day is assumed, the total flow is 0.5 million gal. per day (mgd) and the phosphorus concentration in the wastewater is as follows:

$$\frac{48}{0.5 \times 8.34} = 11.5 \text{ mg/l (as P)}$$

The average total phosphorus concentration in domestic raw wastewater is found to be about 10 mg/l expressed as elemental phosphorus (P).

The above information is provided to serve as a rough guide to engineers and not as a basis of design. Sampling of the wastewater and analyses for phosphorus are recommended in all cases. It may also be necessary to survey the industries that will be discharging wastewater to the sanitary sewer system, to determine their influence on the concentration of phosphorus in the combined wastewater.

Among industrial sources of phosphorus may be potato processing wastes (3), fertilizer manufacturing wastes (4), animal feedlot wastes (5), certain metal finishing wastes (1, 6), flour processing wastes (7), dairy wastes (8, 9), commerical laundry wastes (10), and slaughterhouse wastes (11). The amount of phosphorus may be quite variable depending on the specific industrial plant.

2.2 State Regulatory Agencies

Each state has at least one organization responsible for water pollution prevention, control, and abatement activities. In about half of the 50 states this responsibility is divided between two state agencies. Usually one of these is the state health department and the other an organization charged specifically by statute with the conduct of the state's water pollution control program. Although there has been an intensification of Federal activities in water pollution control over the past twenty years, much of the responsibility for program operations rests with the states. Invariably the appropriate state agencies should be consulted in connection with the planning and conduct of water pollution control programs at cities and other communities. The Environmental Protection Agency regional office personnel work closely with state agencies and can give advice regarding the proper state agency contacts.

2.3 State Standards

As of June 1971, sixteen states had adopted wastewater effluent phosphorus standards. In general, these standards have taken the form of an effluent concentration limit or a requirement for a specified percentage reduction in the phosphorus concentration in the raw wastewater. In most cases, effluent concentration limits range from 0.1 to 2.0 mg/l as P, with many established at 1.0 mg/l. Percentage reduction requirements range from 80 to 95%.

It should be noted that neither an effluent nor a percentage-reduction standard actually limits the phosphorus load in terms of pounds of phosphorus discharged per day. Load (lb/day) is a direct function of both effluent P concentration and daily effluent volume. Assuming the effluent concentration standard is being met, an increase in the daily flow of effluent will produce a proportionate increase in the mass of phosphorus discharged daily to the receiving waters. If the phosphorus load then becomes intolerable, adjustment downward of the effluent concentration standard will be necessary. Similarly, adjustment upward of a percentage-reduction standard will be required if the phosphorus load becomes objectionable as a result of increased wastewater flow or increased P concentration in the raw wastewater or both.

Effluent concentration standards can be derived from receiving water standards as indicated in the following example.

Receiving Water	
Phosphorus concentration upstream	0.01 mg/l

Receiving Water - (continued)	
Maximum allowable phosphorus concentration	0.05 mg/l
Low flow	140 ft^3/sec (90 mgd)
Wastewater	
Average daily flow	4.8 mgd
Effluent concentration standard	X

Mass Balance

$$(0.01)(8.34)(90) + (X)(8.34)(4.8) = (0.05)(8.34)(90+4.8)$$

$$X = 0.8 \text{ mg/l P}$$

Implicit in the above calculation is the assumption that the effluent is mixed thoroughly with the waters of the receiving stream. The low flow selected is not necessarily the low flow of record. It is usually the low flow averaged over a period of 10 days or less and having a return period of once in 10 years.

2.4 Provision for Future Plant Expansion and More Stringent Effluent Requirements

The plant site is an important consideration in the design of treatment works. It is highly desirable that sufficient site area be obtained to permit initial construction of the required system without crowding and to provide space for future expansion. Treatment systems are likely to become larger and more complex in the future rather than less so, and failure to obtain ample site area initially is almost certain to create costly problems later.

Reduction of the allowable effluent phosphorus concentration may be necessary in some states as increases in wastewater flow or phosphorus concentration of the raw wastewater occur. In some instances, merely increasing the chemical dosage may be the only necessary change. On the other hand, it may be necessary to add an additional treatment sequence such as filtration to the existing facilities. In nearly every case, planning for plant expansion or more stringent effluent requirements will result in economy. Such planning should include selection of a plant layout that allows space for addition of facilities without removing existing processes from service.

Other provisions that should be considered are space within chemical buildings for additional coagulant storage and feeding equipment, space for polymer storage and feed equipment, and adequate space for handling the additional solids generated by phosphorus removal processes.

2.5 Flow Equalization

The cyclic nature of wastewater flows, in terms of volume and strength, is well established. While the concept of flow equalization has been employed in the field of water supply, and in the treatment of some industrial wastes, it has not been widely accepted in the municipal pollution control field. Anticipated problems with solids settling, odor, and septicity can be cited as the major factors limiting its use.

The advent of stricter stream standards, which sometimes require removal of contaminants during peak flows, the elimination of plant bypassing, and the increased removal efficiencies that are possible when biological or chemical treatment processes are operated at near steady-state conditions, have all favored the increased use of flow-smoothing, and flow equalization devices ahead of or integral with the design of major treatment works.

There are two major objectives in the design of flow equalization basins. The first of these is simply to dampen the diurnal flow variation that normally exists in typical municipal wastewater collection systems to achieve a constant or nearly constant flow rate through the downstream treatment processes. In this type of system, little consideration is given to controlling the concentration changes that take place during storage. The major design factors are supplying enough air to keep the basin aerobic and to provide adequate mixing to prevent solids deposition.

The second objective of flow equalization is to provide the capacity to distribute shock loads of toxic or process inhibiting substances over a reasonable period of time so as to prevent system failure and to minimize the periodic discharge of harmful contaminants to the receiving stream or surface impoundment. Time-dependent concentration profiles and flow-through curves are normally used to analyze the flow characteristics of these systems and to determine the effect of tank geometry, placement of effluent weirs and mixing regime on changes in contaminant concentrations through the basin.

In all cases the added costs of flow equalization must be measured against the possible reduction in downstream process costs and the increased efficiencies that can be achieved by operating these processes under constant loading conditions.

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Chapter 3

THEORY OF PHOSPHORUS REMOVAL BY CHEMICAL PRECIPITATION

3.1 Forms and Measurement of Phosphorus

Phosphorus is found in wastewater in three principal forms: orthophosphate ion, polyphosphates or condensed phosphates, and organic phosphorus compounds. Orthophosphate was in the past often considered to be PO_4^{3-} and was reported in this way. Actually, there are a number of forms of orthophosphate in equilibrium, with the predominant form changing as pH changes. At the usual pH of municipal wastewater, the predominant form is HPO_4^{2-}. With the present practice of reporting phosphorus as P, one does not usually have to be concerned with the actual form. It is well to remember, however, that the predominant form does change with pH. The removal of phosphorus with lime results primarily from pH increase. Polyphosphates can be looked upon as polymers of phosphoric acid from which water has been removed. Materials of various molecular weight are in widespread use. Complete hydrolysis results in formation of orthophosphate. The chemistry of organic phosphorus compounds is complicated. Their decomposition also leads to orthophosphate.

In raw sewage there are substantial amounts of all three principal phosphorus forms. During biological treatment, significant changes take place. As organic materials are decomposed, their phosphorus content is converted to orthophosphate. On the other hand, inorganic phosphates are utilized in forming biological floc. The polyphosphates are for the most part converted to orthophosphate. The result is that in a well treated secondary effluent, a large fraction of the phosphorus is present as orthophosphate. From the standpoint of removal by precipitation, this is fortunate since the orthophosphate is the easiest form to precipitate.

Because phosphorus will be present in a variety of forms, both organic and inorganic, the only satisfactory measure of treatment plant removal efficiency must be based on total phosphorus entering the plant in the raw wastewater and total phosphorus discharged in the plant effluent. Inasmuch as all colorimetric tests for phosphorus depend upon the formation of an orthophosphate color complex, it is essential in the determination of total phosphorus, to convert any polyphosphates (condensed phosphates) and organic phosphorus compounds present in the sample to the orthophosphate form. This conversion is accomplished by acid digestion of the sample in the presence of a strong oxidizing agent. *Standard Methods*, 13th edition (1), describes several procedures for determining total phosphorus as well as the various forms of phosphorus in wastewater. A suitable procedure for measuring total phosphorus, including the preliminary digestion stage, may be selected from this reference. An automated method is described in *FWPCA Methods for Chemical Analysis of Water and Wastes (2)*.

3.2 Chemistry of Removal by Precipitation

The materials found practical for phosphorus precipitation include the ionic forms of aluminum, iron, and calcium. The calcium is added as lime and the hydroxyl ions also have a role in phosphorus removal. The chemistry of phosphorus removal by interaction with metallic ions is complex. In the interests of brevity and simplicity, it is assumed in the following discussion that phosphorus reacts as the orthophosphate form, PO_4^{3-}, and that the reaction products are as indicated. The polyphosphates and organic phosphorus are also removed, probably by a combination of more complex reactions and sorption on floc particles. The reactions presented and the associated computations are principally for illustrative purposes and do not purport to represent the true reactants, products and reaction mechanisms. Calculated requirements for phosphorus precipitants must be viewed as no more than crude estimates subject to large variations in actual practice because the forms of phosphorus and the precipitates may be substantially different from those assumed.

3.2.1 Aluminum Compounds as Phosphorus Precipitants

Aluminum ions can combine with phosphate ions to form aluminum phosphate as follows:

$$Al^{3+} + PO_4^{3-} \longrightarrow \underline{AlPO_4}$$

The above equation indicates that the mole ratio for $Al:PO_4$ is 1:1. Inasmuch as the mole ratio of $P:PO_4$ is also 1:1, the mole ratio for Al:P is 1:1 or $\frac{Al}{P} = 1$ when both aluminum and phosphorus are expressed in terms of gram-moles or lb-moles. On a weight, rather than a mole basis, this means that 27 lb of Al will react with 95 lb of PO_4 to form 122 lb of $AlPO_4$. Since each 95 lb (1 lb-mole) of PO_4 contains 31 lb (1 lb-mole) of P, the weight relationship between Al and P is 27 lb of Al to 31 lb of P or 0.87 for this reaction.

The principal source of aluminum for use in phosphorus precipitation is "alum", a hydrated aluminum sulfate, having the approximate formula $Al_2(SO_4)_3 \cdot 14H_2O$ (molecular weight of 594). The chemical, which is also known as "filter alum" or "papermaker's alum", averages about 17% soluble aluminum expressed as Al_2O_3 or about 9.1% expressed as Al. Its reaction with PO_4^{3-} may be written as follows:

$$Al_2(SO_4)_3 \cdot 14H_2O + 2PO_4^{3-} \longrightarrow \underline{2AlPO_4} + 3SO_4^{2-} + 14H_2O$$

The sulfate remains in solution as SO_4^{2-}. The above reaction indicates that 1 lb-mole of alum (594 lb) will react with 2 lb-moles (190 lb) of PO_4^{3-} containing 62 lb phosphorus to form 2 lb-moles (244 lb) of $AlPO_4$. The weight ratio of alum to phosphorus is, therefore, 9.6:1. The alum requirement per pound of phosphorus may also be derived from the Al:P mole ratio as follows:

Mole ratio Al:P = 1:1
Therefore weight ratio Al:P = 27:31 = 0.87:1
Alum contains 9.1% Al
Therefore, alum required per lb of P = $\frac{0.87}{0.091}$ = 9.6 lb

Stumm and Morgan (3), state that the solubility of $AlPO_4$ is pH dependent and varies as follows:

pH	Approximate Solubility (mg/l)
5	0.03
6	0.01 (minimum solubility)
7	0.3

The optimum pH for removal of phosphorus probably lies in the range of 5.5 to 6.5, although some removal occurs above pH 6.5.

Addition of alum will lower the pH of wastewater because of neutralization of alkalinity and release of carbon dioxide. The extent of pH reduction will depend principally on the alkalinity of the wastewater. The higher the alkalinity, the less is the reduction in pH for a given alum dosage. Most wastewaters contain sufficient alkalinity so that even large alum dosages will not lower the pH below about 6.0 to 6.5. In exceptional cases of low wastewater alkalinity, pH reduction may be so great that addition of an alkaline substance, such as sodium hydroxide, soda ash, or lime will be required.

Adjustment of the pH downward can be accomplished by the addition of sulfuric acid but this complicates the treatment process and it may be preferable simply to use a higher alum dosage.

Bench, pilot, and full-scale studies have shown that considerably higher than stoichiometric quantities of alum usually are necessary to meet phosphorus removal objectives. A competing reaction, responsible for the pH reduction mentioned above, at least partially accounts for the excess alum requirement. It occurs as follows:

$$Al_2(SO_4)_3 \cdot 14H_2O + 6HCO_3^- \longrightarrow \underline{2Al(OH)_3} + 6CO_2 + 14H_2O + 3SO_4^{2-}$$

The following ratios of alum (9.1%Al) to phosphorus are believed reasonably representative for alum treatment of municipal wastewater (4).

P Reduction Required	Al:P		Alum:P
	Mole Ratio	Weight Ratio	Weight Ratio
75%	1.38:1	1.2:1	13:1
85%	1.72:1	1.5:1	16:1
95%	2.3:1	2.0:1	22:1

To achieve 85% P removal from a wastewater containing 11 mg/l of P the alum dosage needed would be (16)(11) = 176 mg/l or 1470 lb/10^6 gal.

Sodium aluminate will also serve as a source of aluminum for the precipitation of phosphorus. The chemical formula for sodium aluminate is $Na_2Al_2O_4$ or $NaAlO_2$.

One commercial form is the granular trihydrate, which may be written $Na_2O \cdot Al_2O_3 \cdot 3H_2O$ and which contains about 46% Al_2O_3 or 24% Al.

The sodium aluminate - phosphate reaction may be represented as follows:

$$Na_2O \cdot Al_2O_3 + 2PO_4^{3-} + 4H_2O \longrightarrow \underline{2AlPO_4} + 2NaOH + 6OH^-$$

In contrast to alum, which reduces pH, a rise in pH may be expected upon addition of sodium aluminate to wastewater.

The above reaction indicates an Al:P mole ratio of 1:1, an Al:P weight ratio of 0.87:1, and a sodium aluminate to P weight ratio of about 3.6:1.

As with the use of alum, mole and weight ratios somewhat in excess of the theoretical must be anticipated in practice.

3.2.2 Iron Compounds as Phosphorus Precipitants

Both ferrous (Fe^{2+}) and ferric (Fe^{3+}) ions can be used in the precipitation of phosphorus. With Fe^{3+} a reaction can be written similar to that shown earlier for precipitation of aluminum phosphate. A 1:1 mole ratio of $Fe:PO_4$ results. Since the molecular weight of iron is 55.85, the weight ratio of Fe:P is 1.8:1. Just as in the case of aluminum, a larger amount of iron is required in actual situations than the chemistry of the reaction predicts. With Fe^{2+} the situation is more complicated and not fully understood. Ferrous phosphate can be formed. The mole ratio of $Fe:PO_4$ would be 3:2. Experimental results indicate, however, that when Fe^{2+} is used, the mole ratio of Fe:P will be essentially the same as when Fe^{3+} is used.

A number of iron salts are available for use in phosphorus precipitation. These include ferrous sulfate, ferric sulfate, ferric chloride, and pickle liquor. Additional discussion of these is given later. All will lower the pH of wastewater because of neutralization of alkalinity. Pickle liquor contains substantial amounts of free sulfuric acid or hydrochloric acid.

Iron salts are most effective for phosphorus removal at certain pH values. For Fe^{3+} the optimum pH range is 4.5 to 5.0. This is an unrealistically low pH, not attained in most wastewaters. Significant removal of phosphorus can be attained at higher pH. For Fe^{2+} the optimum pH is about 8 and good phosphorus removal can be obtained between 7 and 8. The acidity of ferrous salts, and especially pickle liquor, necessitates addition of lime or sodium hydroxide for good results. Where the water is aerated following Fe^{2+} addition, the use of a base may not be necessary.

3.2.3 Lime as a Phosphorus Precipitant

Calcium ion reacts with phosphate ion in the presence of hydroxyl ion to form hydroxyapatite. This material has a variable composition, but an approximate equation for its formation can be written as follows, assuming in this case that the phosphate present is HPO_4^{2-}:

$$3\ HPO_4^{2-} + 5Ca^{2+} + 4OH^- \longrightarrow \underline{Ca_5(OH)(PO_4)_3} + 3\ H_2O$$

The reaction is pH dependent. The solubility of hydroxyapatite is so low, however, that even at a pH as low as 9.0, a large fraction of the phosphorus can be removed. In lime treatment of wastewater, the operating pH may be predicated on the ability to obtain good suspended solids removal rather than on phosphorus removal.

Although it is possible to calculate an approximate lime dose for phosphorus removal, this is generally not necessary. In contrast to iron and aluminum salts, the lime dose is largely determined by other reactions that take place when the pH of wastewater is raised. Some of these reactions are discussed later. Only in waters of very low bicarbonate alkalinity would the phosphate precipitation reaction consume a large fraction of the lime added.

3.3 References – Chapter 3

1. *Standard Methods for the Examination of Water and Wastewater,* 13th Edition, American Public Health Assoc., New York (1971).
2. *FWPCA Methods for Chemical Analysis of Water and Wastes.* Federal Water Pollution Control Administration, U.S. Department of the Interior (Nov, 1969).
3. Stumm, W. and Morgan, J. J., *Aquatic Chemistry,* p 521 Wiley-Interscience, John Wiley & Sons, Inc., New York (1970).
4. Allied Chemical Corporation, Industrial Chemicals Division, "Use of Aluminum Sulfate for Phosphorus Reduction in Wastewaters", Unpublished paper.

Chapter 4

PHOSPHORUS REMOVAL BY MINERAL ADDITION BEFORE THE PRIMARY SETTLER

4.1 General Considerations

Removal of phosphorus from wastewater can be accomplished by modifying the conventional primary treatment process to include chemical precipitation with aluminum or iron salts (1,2). The advantages of phosphorus removal during primary treatment include flexibility in chemical feeding, adequate reaction times and mixing conditions, flocculation and removal of more suspended solids and suspended BOD in the primary settler, and reduced loading of suspended solids and BOD to secondary treatment processes. Significant increases in removals of both suspended solids and BOD can occur concurrently with phosphorus removal. Phosphorus removal by chemical addition is amenable to automatic control using process instrumentation to measure waste flow and/or phosphorus concentration.

The disadvantage of addition of metal salts during primary treatment alone is that a significant amount of the phosphorus may not be in the ortho form and may not be easily precipitated. Higher phosphorus removals can be obtained with chemical addition beyond primary treatment and with chemical additions at several points.

The available precipitation-flocculation processes (3,4,5,6,7,8) include addition of ferrous or ferric chloride, ferrous or ferric sulfate, aluminum sulfate, or sodium aluminate, followed by flocculation using an anionic polymer to enhance solids separation. A strong base is often added between additions of ferrous iron and polymer to counteract the depression of pH by the ferrous compounds.

In order to provide effective chemical additions to primary treatment units, the following sequence is recommended:

1. Salt addition to raw sewage with thorough mixing. If a base is to be used, it should be added at least 10 seconds later.
2. Reaction for at least five minutes.
3. Addition of anionic organic polymer followed by flash mixing for 20 to 60 seconds.
4. Mechanical or air flocculation for 1 to 5 minutes.
5. Gentle delivery of flocculated wastewater to sedimentation units.

Mixing may be carried out in specially designed basins or advantage may be taken of existing high turbulence areas such as in a Parshall flume. Flocculation may similarly be carried out in specially constructed equipment or may be allowed to take place within some part of an existing system. Schematic diagrams of 2 plants, which demonstrate how existing equipment can be used with mineral addition, are shown in Figures 4-1 and 4-2.

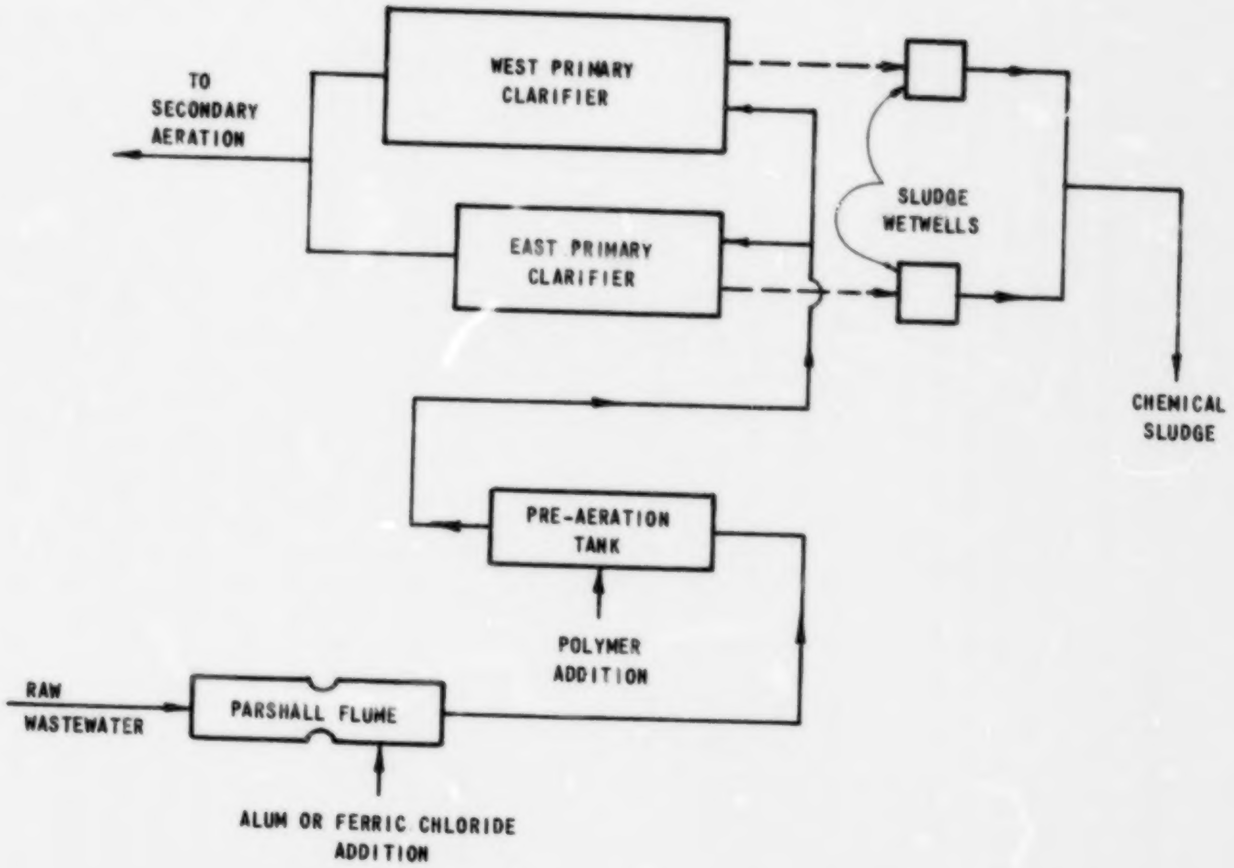
TO SECONDARY AERATION
WEST PRIMARY CLARIFIER
EAST PRIMARY CLARIFIER
SLUDGE WETWELLS
CHEMICAL SLUDGE
PRE-AERATION TANK
POLYMER ADDITION
RAW WASTEWATER
PARSHALL FLUME
ALUM OR FERRIC CHLORIDE ADDITION

FIGURE 4-1 LEBANON, OHIO PRIMARY TREATMENT AND CHEMICAL FEEDING SCHEMATIC

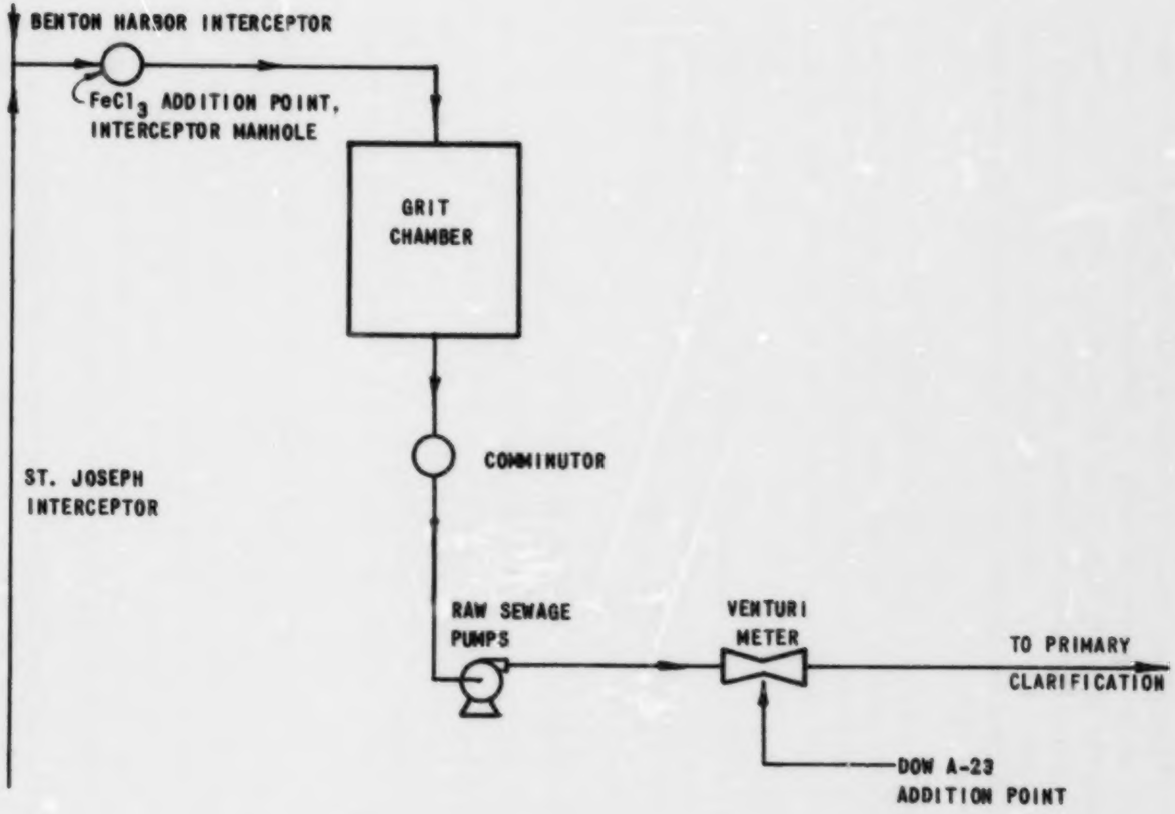
BENTON HARBOR INTERCEPTOR
$FeCl_3$ ADDITION POINT, INTERCEPTOR MANHOLE
GRIT CHAMBER
COMMINUTOR
ST. JOSEPH INTERCEPTOR
RAW SEWAGE PUMPS
VENTURI METER
TO PRIMARY CLARIFICATION
DOW A-23 ADDITION POINT

FIGURE 4-2 SCHEMATIC OF PRETREATMENT OF PLANT FLOW AT BENTON HARBOR - ST. JOSEPH WASTEWATER DISPOSAL PLANT

The points of chemical addition will vary depending on the particular situation. It is important, however, that addition points be downstream of return streams, such as digester supernatant, if a high degree of phosphorus removal is required.

Optimum conditions for phosphorus removal processes are best determined in two steps: a laboratory feasibility study followed preferably by full-scale plant trials if the plant exists, or by pilot trials. The feasibility study should be a short-term laboratory evaluation of various chemical systems to obtain preliminary estimates of chemicals required to meet established phosphorus removal standards. During the plant trials, chemicals and equipment should be supplied for full-scale operation over a period of 30 days or longer, to obtain detailed design data.

Chemical removal of phosphorus from wastewaters during primary treatment is a practical and controllable process which can result in low discharges of phosphorus from existing primary treatment plants and secondary treatment plants of either the trickling filter or activated sludge type. Additional costs for chemicals and attendant mixing, dispersing, and monitoring equipment are moderate.

Data and correlations given in this chapter will provide guidance and permit making design estimates. However, it would be risky to eliminate on-site experimental studies because of variations in equipment design and operation, and in wastewater characteristics.

4.2 Phosphorus Removal Data for Alum Addition

Alum has been used as a source of aluminum for phosphorus removal in raw wastewater. Sodium aluminate can also be used if the natural alkalinity of the raw wastewater is too low to prevent excessive reduction of pH by alum. However, no investigations using aluminate have been reported. Tests have been conducted on laboratory, pilot, and plant scales to determine the feasibility of alum addition to raw wastewater.

Bench-scale tests of alum addition were conducted at Springfield, Ohio (9) and Two Rivers, Wisconsin (10). All bench testing was conducted in the following manner. Samples of wastewater were stirred on a six-place stirrer for approximately 2 minutes at 100 rpm, followed by 18 minutes of gentle stirring (40 to 50 rpm). Alum additions were made during the rapid mix period. When polymers were used, they were added approximately 2 minutes after the alum addition. Treated samples were allowed to settle for 30 minutes before total phosphorus analyses were made.

The respective states have required phosphorus removals of 80% at Springfield and 85% at Two Rivers. At Springfield, an average Al:P weight ratio of 1.9:1 (molar ratio of 2.2:1) was required to reduce the average influent total phosphorus of 5 mg/l by 80%. Raw wastewater at Two Rivers required an average Al:P weight ratio of 0.93:1 (molar ratio of 1.07:1) to reduce the average influent phosphorus of 20.1 mg/l by 85%. These test data indicate that as phosphorus concentration increases, the Al:P ratio necessary to obtain a specific per cent removal decreases. Individual tests at Two Rivers, for a broad range of influent phosphorus concentrations, also showed the same variation. Influent total phosphorus ranged from 7.6 to 28.5 mg/l, and the Al:P weight ratios necessary to remove 85% ranged from 2.0:1 to 0.58:1, respectively.

A study of alum addition for phosphorus removal was conducted at Lebanon, Ohio on both pilot and plant scales (11) to determine whether chemical treatment of raw wastewater could be used successfully to precede activated carbon adsorption for organic removal.

A 15 gpm pilot plant was utilized to determine the most effective dosages of alum to be studied in the full-scale tests. Data from the pilot plant indicated that liquid alum dosages of 150, 250, and 350 mg/l, as $Al_2(SO_4)_3 \cdot 14H_2O$, and anionic polymer dosages of 0.25 and 0.5 mg/l should be evaluated. Only the Parshall flume, preaeration basin, and primary clarifiers of the 1.5 mgd Lebanon plant were used in this study. Alum was fed to the raw wastewater at the Parshall flume to assure complete mixing. The preaeration basin was used for flocculation, with polymer being added toward the end of the basin. The primary basins served as settlers for the precipitated phosphorus. Figure 4-1 shows the chemical addition arrangement at Lebanon which was used for both alum and iron studies. The iron experiments will be covered later in this chapter.

Each dosage was evaluated for five days. A polymer dose of 0.5 mg/l was found to be satisfactory for improving floc formation and settling. An alum dose of 250 mg/l, Al:P weight ratio of 3.4:1 (molar ratio of 3.9:1), removed 93% of the phosphorus in this high alkalinity (400 mg/l as $CaCO_3$) wastewater. Effluent phosphorus concentration did not exceed 0.7 mg/l and averaged 0.5 mg/l. The alum dose of 150 mg/l, Al:P weight ratio of 1.5:1 (molar ratio of 1.7:1), removed only 74% of the phosphorus; while the alum dose of 350 mg/l, Al:P weight ratio of 3.9:1 (molar ratio of 4.5:1), reduced the phosphorus by 94%. The alum dose needed to meet the State removal requirement of 80% would fall between the 150 and 250 mg/l dosages studied.

Little data are available for sludges resulting from alum addition to primary settlers. At Lebanon, Ohio, the primary sludge produced by alum addition received lime treatment before final disposal as landfill or on farm land (12). Before lime treatment, the sludge was similar in character to biological sludge and had a solids concentration of from 2 to 3%. It had very poor settling and thickening characteristics and was malodorous.

Lime treatment consisted of mixing the sludge with a 25% lime slurry until a pH of 11.5 was reached. The average lime requirement was 440 lb of lime as $Ca(OH)_2$ per ton of dry solids. A contact time of 30 minutes was maintained before the sludge was allowed to flow to sand beds for drying. Seventy per cent of the total volume of sludge applied to the drying beds was removed as drainage in the first 3 days, and the sludge cake could be lifted off the beds in 3 weeks. Elimination of odors and complete pathogen kill was accomplished with lime treatment. Lime costs at $18/ton for $Ca(OH)_2$, and an application rate of 440 lb/ton of dry solids, resulted in a chemical cost for sludge handling of $3.96/ton of dry sludge solids.

4.3 Phosphorus Removal Data for Iron Addition

A number of studies have been conducted using the addition of iron to raw wastewater for phosphorus removal. These investigations have ranged from laboratory jar testing to full plant-scale work and have included both ferrous and ferric salts as the source of iron. Average values for doses, phosphorus removals, and references to the studies are listed in Table 4-1.

These studies include wastewaters with phosphorus concentrations of 6 to 9 mg/1, so a range of phosphorus removals would be expected. In addition, there are many other significant variations in wastewater character from location to location. The results of the studies allow bracketing the phosphorus removals obtained and enable drawing conclusions based on special features of individual studies.

With primary treatment alone, phosphorus removals varied from 60 to 91%. In secondary plants, iron addition before the primary resulted in phosphorus removals from 75 to 93% over the entire plant. Iron doses varied between 10 and 90 mg/1, as Fe. The Escanaba study showed that base and polymer addition could halve the iron dose required to obtain the same phosphorus removal achieved with iron alone. The Lebanon study showed that doubling the iron dose (from 45 to 90 mg/1) with the same polymer addition raised phosphorus removal from 68 to 91%. At Wyoming, Michigan, the same phosphorus removal was obtained with equal doses of either Fe^{2+} or Fe^{3+}. Polymer doses in all of the iron studies ranged from 0.26 to 0.7 mg/1.

Iron carry-over occurred at Mentor, where $FeCl_2$ addition alone resulted in an effluent Fe concentration of 42.5 mg/1. Lime addition reduced the iron leakage to an effluent Fe concentration of 11.6 mg/1. Polymers were ineffective in reducing iron leakage.

Increased suspended solids and BOD removals were also obtained with iron additions. At Grayling, Michigan, suspended solids removal was increased by 27% to an average of 78%. At Benton Harbor, Michigan, where waste activated sludge is returned to the primary, iron addition gave a 3.0% increase in solids removal over the entire treatment plant. An example of increased BOD removal is seen in the Grayling, Michigan study. Removal of BOD in the primary was increased by 45% after iron addition. This resulted in a total BOD removal of 58% in the primary.

Typical values for primary settler detention times were 2.3 hours at Mentor and 2.25 hours at Benton Harbor. At Mentor, air at a rate of 42.5 to 85 ft^3/minute was used for flocculation in the mixing area of the primary tank.

Although lab test procedures should be adjusted to best approximate individual plant conditions, outlines of procedures used at two locations may be useful as guidelines. Jar tests were conducted at Mentor in the following manner. Iron, lime, and polymer

Table 4-1

STUDIES OF IRON ADDITION BEFORE THE PRIMARY SETTLER FOR PHOSPHORUS REMOVAL

Location	Type of Study	Iron Form	Fe Dose (mg/l) and/or Fe:P (by weight)	Base (mg/l)	Polymer (mg/l)	% P Removal in Primary
Grayling, Michigan (13)	plant scale	$FeCl_2$	15-25	30-50 (NaOH)	0.3 (Dow A23, anionic)	60-80
Lake Odessa, Michigan (14)	plant scale	$FeCl_2$	15-25	30-50 (NaOH)	0.3 (Dow A23, anionic)	75*-93*
Bay City, Michigan (15)	lab	$FeCl_2$	10, 1.8:1	30 (CaO)	0.5 (Dow A23, anionic)	80
Escanaba, Michigan (16)	lab	Fe^{3+}	2.2:1	0	0	80
		Fe^{3+}	1.1:1	40 (CaO)	0.35 (Dow A23, anionic)	80
Wyoming, Michigan (17,18,19)	lab	Fe^{2+}	10	30 (CaO)	0.7 (Dow A23, anionic)	80
		Fe^{3+}	10		0.4 (Dow A23, anionic)	80

Table 4-1 (Continued)

Location	Type of Study	Iron Form	Fe Dose (mg/l) and/or Fe:P (by weight)	Base (mg/l)	Polymer (mg/l)	% P Removal in Primary
Benton Harbor-St. Joseph, Michigan (20)	lab	$FeCl_3$	20		0.3 (Dow A23, anionic)	optimum
	5.5 mgd** (activated sludge return to primary)	$FeCl_3$	21		0.26 (Dow A23, anionic)	89.7*
	5.5 mgd** (no sludge return to primary)	$FeCl_3$	21		0.26 (Dow A23, anionic)	65.3 90.9*
Cleveland, Ohio (21)	35 mgd-Imhoff tanks as primaries	Fe^{3+}	20		0.5 (Dow A23, anionic)	78***
Warren, Michigan (22)	36 mgd	$FeCl_3$	15			80*
Lebanon, Ohio (11)	1.01 mgd	$FeCl_3$	45, 1.76:1		0.5 (Atlas Altasep 2A2, anionic)	68
	0.86 mgd	$FeCl_3$	90, 3.46:1		0.5 (Atlas Altasep 2A2, anionic)	91

Table 4-1 (Continued)

Location	Type of Study	Iron Form	Fe Dose (mg/l) and/or Fe:P (by weight)	Base (mg/l)	Polymer (mg/l)	% P Removal in Primary
Mentor, Ohio (23,24)	lab	$FeCl_2$ (waste pickle)	60	80 (CaO)		>80
	4 mgd	$FeCl_2$ (waste pickle)	3.1:1	CaO:Fe, by weight was 1.9:1, to adjust pH to 7.5-8.0		80

*Secondary treatment provided. Removal value is for entire plant.

**Operated at 5.5 mgd. Design capacities were 8 mgd for primary and 4 mgd for secondary, "Kraus Process", treatment.

***% reduction in effluent P.

were added one minute apart to 1,000 ml of raw wastewater at a stirrer speed of 100 rpm. The chemicals and raw wastewater were then mixed for five minutes at 20 rpm, and finally the mixture was allowed to settle quiescently for 1/2 hour before a portion of the supernatant was pipetted off for analysis. At Bay City, Michigan, the optimum time intervals in the chemical addition sequence were found to be 10 seconds between the iron and lime addition followed by 5 minutes mixing before the addition of Dow Purifloc A23 polymer.

Information related to iron sludges and sludge handling was obtained at Lake Odessa and Grayling, Michigan, and Cleveland, Lebanon and Mentor, Ohio. The sludge resulting from the addition of iron for phosphorus removal can be handled by anaerobic digestion, vacuum filtration, incineration, a combination of these three, or lime treatment. Each of the first three methods has been used during plant-scale studies of iron addition. The last method was used on a research study during a plant-scale addition of iron.

At Lake Odessa, anaerobic sludge digestion is practiced and digester supernatant was returned to the primary unit during the test period. The total phosphorus concentration in the supernatant decreased during the study from an initial concentration of 100 to 200 mg/l to 23 mg/l during latter stages of the study period, and it was concluded that phosphorus release was not occurring during digestion. A crystalline phosphate precipitate found in the raw sludge from Grayling and the digested sludge from Lake Odessa was identified by X-ray diffraction techniques as vivianite, $Fe_3(PO_4)_2 \cdot 8H_2O$. Analyses by X-ray diffraction also showed that the two sludges contained 20 to 24% iron.

The chemical sludge resulting from iron addition at Mentor, Ohio (23, 24), was digested anaerobically and dewatered on vacuum filters. The volume of sludge produced during the iron and lime treatment was never more than double that of periods without chemical treatment. Iron-lime primary sludge averaged 8.5 to 10.0% solids in November and December, 1969. The anaerobic digesters operated with normal gas production and pH. The larger volumes of sludge shortened digester residence time but stabilization did occur. Typical concentrations of iron and phosphorus in the various sludges are given below in Table 4-2.

Table 4-2

TYPICAL IRON AND PHOSPHORUS ANALYSIS
DURING SLUDGE HANDLING AT MENTOR, OHIO

	Primary Sludge Analysis		Digester
	Raw (mg/l)	Digested (mg/l)	Supernatant (mg/l)
Total Iron	3750	1020	3950
Soluble Iron	105	26	trace
Total Phosphorus (as P)	1140	393	1240
Soluble Phosphorus	81.5	32.7	6.6

At Mentor the iron-lime digested sludge had good dewatering characteristics. The digested sludge with 10% solids was dewatered on a vacuum filter to an average of 20% solids.

At Cleveland, Ohio (21) anaerobic digestion, vacuum filtration, and incineration were utilized to dispose of the 28 tons/day of solids produced by $FeCl_3$ addition to the primary units. Polymer conditioning of both raw and digested sludge resulted in filter yield rates of up to 10 lb/ft^2/hr. No problems were encountered in incinerating the cake.

Lime treatment of ferric chloride sludge at Lebanon, Ohio (12) was designed on the basis of 8,700 gpd of sludge. Lime addition to the sludge to a pH of 11.5 required 244 lb of lime, as $Ca(OH)_2$, per ton of dry sludge solids. After 30 minutes of contact at this pH, the sludge was pumped to drying beds. Samples of limed and unlimed sludge were analyzed, and corresponding sludge characteristics are given in Table 4-3. The analyses show that lime treatment increased sludge solids and filter yields but had little or no effect on filter cake moisture content.

Table 4-3

SLUDGE FROM FERRIC CHLORIDE TREATMENT
OF RAW SEWAGE AT LEBANON, OHIO

	Before Lime	After Lime
Sludge Solids, %	1.9	2.4
Volatile Portion, %	70	57
Dissolved Solids, %	0.17	0.17
Vacuum Filter Yield*		
90 mg/l $FeCl_3$ dose	1.06	1.57
45 mg/l $FeCl_3$ dose	1.57	2.40
Filter Cake Moisture Content**		
90 mg/l $FeCl_3$ dose	4.10	4.28
45 mg/l $FeCl_3$ dose	3.75	3.75

*lb of sludge solids/ft^2/hour
**lb of water/lb of sludge solids

4.4 Overall Performance of Treatment Plants with Mineral Addition to the Primary Settler

4.4.1 Primary Treatment Alone

The relative efficiencies of primary and secondary treatment with and without primary phosphorus removal are summarized in Table 4-4. Phosphorus removal without chemical precipitation in the primary clarifiers ranges from only 10 to 20% through a

Table 4-4

EFFECTIVENESS OF PRIMARY AND SECONDARY TREATMENT WITH AND WITHOUT MINERAL ADDITION FOR PHOSPHORUS REMOVAL

	Phosphorus Removal (%)		Suspended Solids Removal (%)		BOD Removal (%)	
	Without	With	Without	With	Without	With
Primary Treatment	5-10	70-90	40-70	60-75	25-40	40-50
Secondary Treatment						
Trickling Filter	10-20	80-95	70-92	85-95	80-90	80-95
Activated Sludge	10-20	80-95	85-95	85-95	85-95	85-95

conventional secondary treatment plant. At least one half or more of this removal is due to sedimentation of the insoluble phosphorus fraction in the raw wastewater during primary clarification. The addition of the precipitation-flocculation process in primary clarification will result in 70 to 90% removal of total phosphorus in the primary. These removal levels can be expected only in well designed primary clarifiers that are not hydraulically overloaded. Suspended solids removals in the primaries will be increased from 40 to 70% to 60 to 75%. This increase occurs despite the additional solids generated by phosphorus removal. BOD removals in the primaries will be increased from 25 to 40% to 40 to 50%. The additional removal is primarily due to the suspended BOD fraction.

4.4.2 Effect on Secondary Treatment

Additional phosphorus removal through either trickling filter or activated sludge units following chemical precipitation in the primaries can amount to 10 to 15%. This removal is attributed to biological flocculation. Split addition of metal in both the primary and secondary portions of the plant should be considered if exceptionally high levels (>90%) of total phosphorus removal are desired. Addition points for the secondary metal feed at or near the effluent end of the biological treatment unit should be included in the design.

Increased capture of suspended solids in the primaries results in reduced loading to the secondary units. Although the solids entering the secondary are reduced, the liquid volumes remain the same and, therefore, normal detention times should be used. Essentially the same amount of soluble BOD must be insolubilized and removed. The additional removal of suspended solids and BOD across the primaries will reduce the amount of waste activated sludge to be handled. Savings in plant operating costs may result from reduced air requirements for aeration, decreased chlorine demand, and improved sludge filterability (25).

4.5 Sludge Handling Requirements

Application of phosphorus removal during primary treatment will result in increased amounts of primary sludge because of the greater capture of suspended solids and the precipitated metal phosphate and hydroxide solids. Consequently, the amount of secondary sludge will generally decrease, and the ratio of primary to secondary sludge will increase.

Sludge conditioning characteristics will also be changed. Instead of requiring cationic polymers or ferric chloride and lime for conditioning, sludges resulting from mineral addition to the primary settler will probably best be conditioned with an anionic polymer. Filter yields should remain about the same as those for sludges from conventional treatment without mineral addition. Addition of polymer to the raw wastewater may reduce the polymer demand for conditioning the sludge.

Anaerobic digestion will be unaffected by the addition of chemically precipitated sludge. The higher digester loadings resulting from additional sludge production will not normally be detrimental to operation unless an organic overloading condition exists. Release of soluble phosphorus from the sludge during digestion is considered to be minimal. In situations requiring exceptionally high degrees of total phosphorus removal (>90%), it may be desirable to treat sidestreams such as digester supernatant and filtrate separately with metal salts before recycle to the head of the plant.

An example of estimation of the quantities of sludge and its distribution in a hypothetical activated sludge plant are given in Table 4-5. Relative concentrations of suspended solids, BOD, and total phosphorus are based on a hypothetical wastewater with process efficiencies in the range of those given in Table 4-4.

Table 4-5

ESTIMATION OF SLUDGE PRODUCTION

Assumed Conditions	Conventional	Chemical
Suspended solids, mg/l (% removed)		
Raw	300	300
Primary	150 (50)	75 (75)
Secondary	30 (90)	15 (95)
Soluble BOD, mg/l (% removed)		
Raw	150	150
Primary	150 (0)	150 (0)
Secondary	30 (80)	10 (93)
Total phosphorus, mg P/l (% removed)		
Raw	10	10
Primary	9 (10)	2 (80)
Secondary	8 (20)	1 (90)

Chemical Dose = 25 mg Fe/l = 73 mg $FeCl_3$/l

Chemical Precipitation

$$\begin{array}{lclcl} Fe^{3+} & + & PO_4^{3-} & \longrightarrow & \underline{FePO_4} \\ 16.2\text{ mg/l} & & 9.0\text{ mg/l P} & & 44\text{ mg/l} \end{array}$$

$$\begin{array}{lclcl} Fe^{3+} & + & 3OH & \longrightarrow & \underline{Fe(OH)_3} \\ 8.8\text{ mg/l} & & 8.1\text{ mg/l} & & 17\text{ mg/l} \end{array}$$

73 mg $FeCl_3$/l gives 44 mg $FePO_4$/l + 17 mg $Fe(OH)_3$/l
= 61 mg precipitated Fe solids/l

1 lb $FeCl_3$ gives 0.60 lb $FePO_4$ + 0.23 lb $Fe(OH)_3$
= 0.83 lb precipitated Fe solids

Table 4-5 (Continued)

Sludge Balance	Conventional	Chemical
Initial suspended solids captured, lb/10^6 gal.		
Primary	1250	1875
Secondary	1000	500
Overall	2250	2375
BOD converted solids, lb/10^6 gal.		
Primary	0	0
Secondary	500	583
Overall	500	583
Precipitated iron solids, lb/10^6 gal.		
Primary	0	448
Secondary	0	0
Overall	0	448
Total solids, lb/10^6 gal.		
Primary	1250	2323
Secondary	1500	1083
Overall	2750	3406
Weight ratio of primary sludge to secondary sludge	0.83	2.15
% Change: chemical over conventional		
Primary sludge		+ 86
Secondary sludge		- 28
Total sludge		+ 24

The relative amounts of $FePO_4$ and $Fe(OH)_3$ are estimated from simplified stoichiometry. Secondary or biological sludge quantities have been estimated on the basis that 0.5 lb of solids is produced per lb of soluble BOD applied.

In comparing a conventional secondary treatment plant with the same plant having chemical treatment in the primary for phosphorus removal, the addition of chemical treatment results in better solids capture, improved BOD conversion, and greatly increased phosphorus removal. The calculated ratio of primary to secondary sludge is more than double that of conventional treatment, and in actual operation would result in improved sludge conditioning. The addition of chemical treatment also reduces the load to the secondary, which should result in some improvements in actual operation. Chemical treatment increases the total sludge mass only 24%, of which about 16% is due to formation of the insoluble iron-phosphate-hydroxide sludge.

4.6 Dosage Selection and Correlation

The material presented in Chapters 6 and 7 on choice of chemicals and dosage in trickling filter plants and activated sludge plants generally applies also to mineral addition in primary plants. Therefore, the reader is referred to these chapters for discussions of requirements for influent phosphorus data, procedures for obtaining dosage design data, calculations of daily doses, and scheduling of dose rate changes to match phosphorus variation. The reader is also referred to Chapter 10 for general information on various chemicals. Material presented here will be limited to comments and/or data on dosage selection, dosage correlation, and influent phosphorus variations.

It is generally desirable to plot or correlate test results on a simple basis to make them more useful. A reasonably good correlation can be obtained by plotting the log of the fraction of soluble phosphorus remaining as a function of the weight ratio of metal to soluble phosphorus. Typical of such plots is Figure 4-3, where both the fraction of soluble phosphorus remaining and the percent removed are plotted on the log scale. The data in Figure 4-3 are from several ferric chloride phosphorus removal studies conducted in the Great Lakes region.

Since there is usually considerable scatter in the data, such correlations should be used very conservatively as a guide for predicting doses. On-site tests under actual plant conditions are always preferred as the basis for such a plot. The required ratio of metal ion to influent soluble phosphorus for a specified phosphorus removal can be obtained from such a plot. This ratio is multiplied by the soluble influent phosphorus concentration in mg/l to determine the required chemical dose in mg/l.

As will be discussed in Chapter 6, the dose rate should be varied during the day to match more closely the varying influent phosphorus concentration. The need for doing this is illustrated by Figure 4-4 where some typical data on the diurnal variation of ortho phosphorus in plant influent are presented.

4.7 Plant Designs for Phosphorus Removal with Iron Addition

The following three designs for proposed expansions of wastewater treatment facilities, at Escanaba, Bay City, and Wyoming, Michigan, include iron addition for phosphorus removal. All three were prepared by Black & Veatch, Consulting Engineers, and are designed for iron addition to the primary basin. Escanaba and Wyoming will use ferric chloride while Bay City will use ferrous chloride.

4.7.1 Escanaba, Michigan

A complete-mix activated sludge plant at Escanaba, Michigan is being constructed to replace a combination trickling filter-conventional activated sludge plant. Provisions have been made in the design to meet the State's requirement of 80% phosphorus removal. Effluent from the plant will flow to Little Bay DeNoc of Lake Michigan.

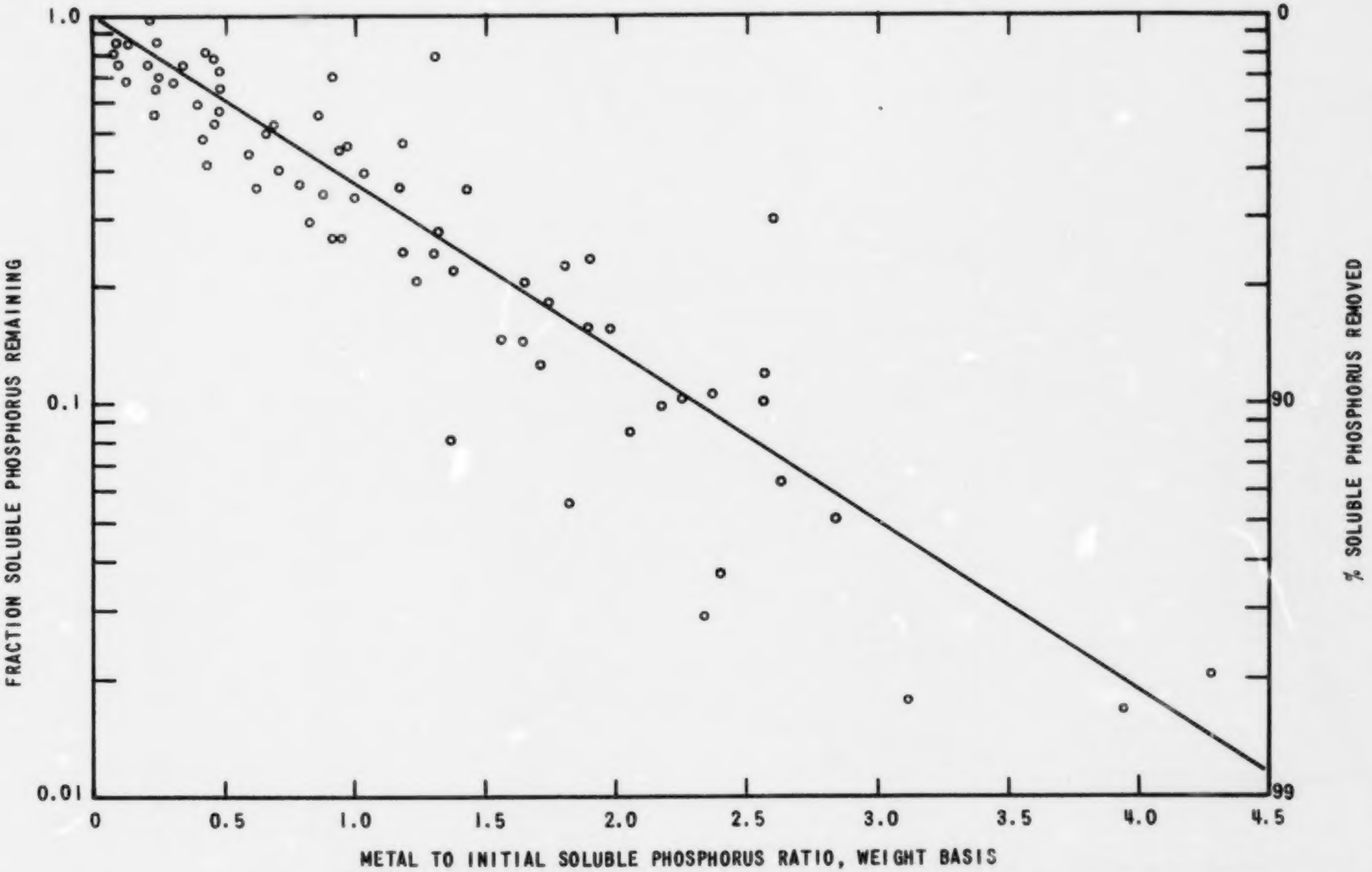

FIGURE 4-3 SOLUBLE PHOSPHORUS REMOVAL BY FERRIC CHLORIDE ADDITION

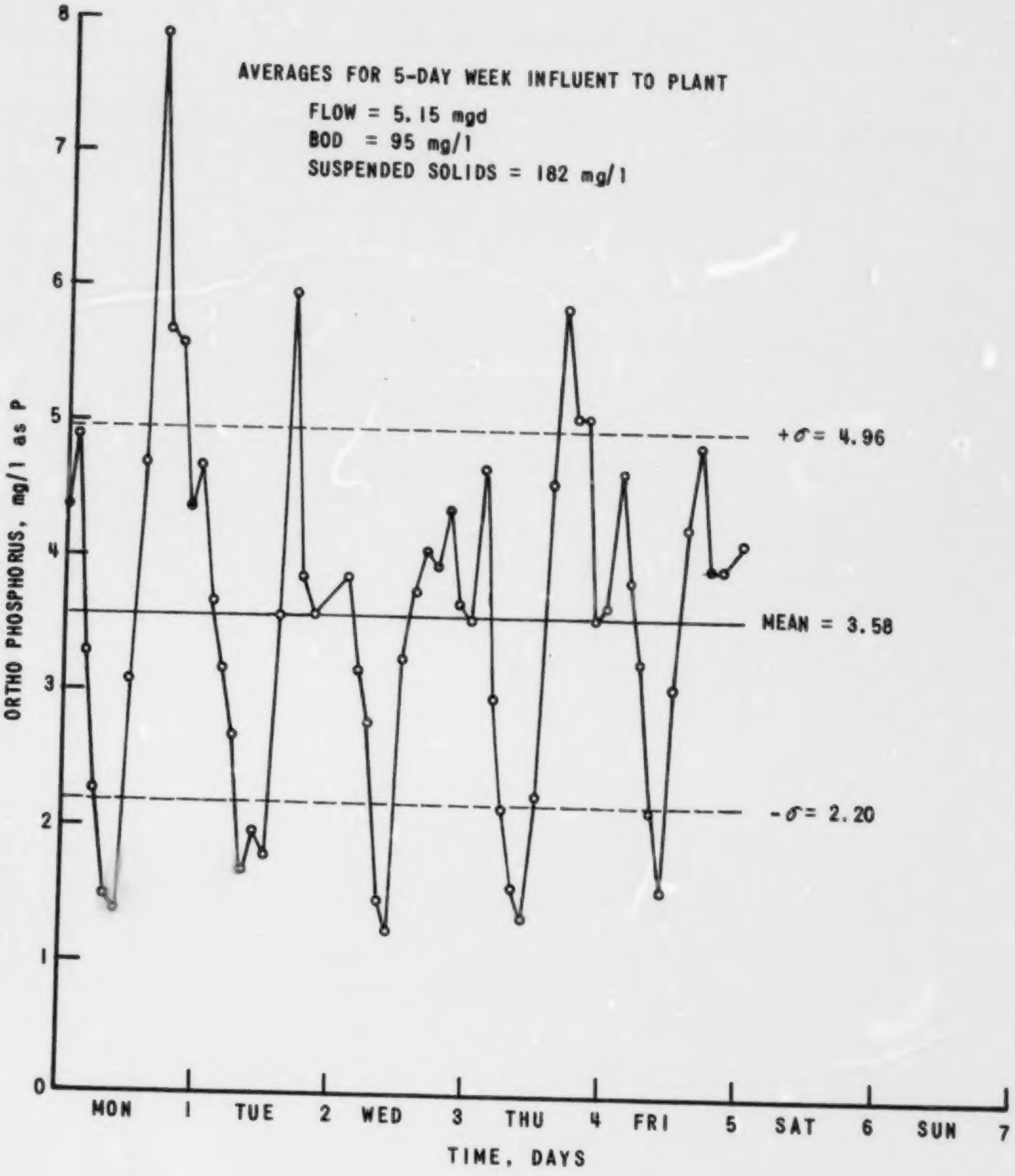

FIGURE 4-4 DIURNAL VARIATION OF ORTHO PHOSPHORUS

Phosphorus removal in the new plant will be accomplished by precipitation with $FeCl_3$ in the primary basin, using polymer to enhance settleability of the floc. Ferric chloride will normally be fed to the wastewater in a new rapid mix basin which will have a 3.4 minute detention time at the average design flow of 2.2 mgd. Alternate iron addition points will be located at the raw wastewater pump station and the Parshall flume. Polymer will normally be added toward the effluent end of an air flocculation basin, having a 65 minute detention at design flow. Polymer may also be added at any of the points of iron addition. Phosphorus precipitation will occur in the four primary basins, each having a detention time of 3 hours at the design flow.

The storage facilities for $FeCl_3$ are designed to provide a 30 day supply at a feed rate of 27 mg/l Fe^{3+} to a plant flow of 2.2 mgd. These facilities consist of two 4,000 gal. rubber lined steel tanks, 8 ft in diameter and 11 ft high. It is anticipated that $FeCl_3$ will be stored and fed at the shipping strength; 35 to 45% solution. However, a metered water supply will be connected to each storage tank to allow dilution, if necessary. Two metering pumps will be furnished and piped to take suction from either storage tank. Either pump will be capable of delivering $FeCl_3$ over a range of from 1.6 to 40 gal./hour. This will allow a feed rate range of from 10 mg/l (as Fe) of 45% solution at a minimum plant flow of 1.0 mgd to 30 mg/l (as Fe) of 35% solution at a peak hourly plant flow of 5.0 mgd. The storage tank and metering feed pump area will be isolated to trap any leakage.

Bagged polymer storage, solution equipment, and metering pumps will be located near the $FeCl_3$ facilities. Polymer solution will be made up in pre-selected concentrations by an automatic volumetric batching system. A metering pump will deliver polymer from a holding tank to any of three application points. This pump is sized for chemical dosages ranging from 0.20 mg/l at 1.0 mgd to 0.30 mg/l at 5.0 mgd. Due to the high viscosity, even of very dilute solutions, polymer (Dow A23) will be fed by a positive displacement rotary gear pump. The feed range for a 0.5% solution will be 5 to 125 gal./hour.

Sludge from the primary basins will be pumped to the existing anaerobic digesters, having a total capacity of 77,700 ft^3. Digested sludge will be pumped to drying beds with a total area of 37,090 ft^2. Waste activated sludge will be digested aerobically.

4.7.2 Bay City, Michigan

The existing wastewater plant at Bay City provides only primary treatment before discharge of the wastewater to Lake Huron through the Saginaw River. The State of Michigan is requiring 80% phosphorus removal and improved organic removal. Iron addition will be used for phosphorus removal and trickling filters will be provided for organic removal. The expanded facilities are now under construction.

The phosphorus removal process design is based on studies conducted by the Dow Chemical Company. These studies indicated that the most efficient method of removing phosphorus would consist of addition of $FeCl_2$, a base (NaOH), and an anionic polymer (Dow A23) to the raw wastewater. Although Dow's recommendations for Bay City's raw wastewater indicated ferrous chloride to be the best iron salt, ferrous sulfate or ferric chloride may also be used as alternatives.

Ferrous chloride will be furnished as a liquor containing a minimum concentration of 22% $FeCl_2$ by weight. The liquor will be shipped in 4,000 to 5,000 gal. tank trucks. Two 20,450 gal. storage tanks will provide a minimum of 2 weeks supply of liquid $FeCl_2$. Feeding of ferrous chloride will be automatically paced with plant flow at present, however, instrumentation is specified such that future control of the iron feed rate in proportion to phosphate concentration is possible if a continuously monitoring phosphate analyzer is added. Positive displacement metering pumps will be provided to feed the liquid $FeCl_2$ into the wastewater at the upstream end of the Parshall flume at the plant inlet.

Sodium hydroxide will be obtained as a 50% caustic soda solution in 3,000 to 4,000 gal. tank trucks. Two 6,450 gal. storage tanks will provide over a month's supply of 50% NaOH, depending on the plant flow and alkalinity of the wastewater. Because 20% caustic solution has a much lower temperature of crystallization, two 1,200 gal. day tanks equipped with mixers will be provided for diluting the 50% solution to 20% before it is pumped from a heated chemical building to the treatment plant. A centrifugal pump will transfer 50% solution to the day tanks when the storage tank levels are low. Positive displacement metering pumps will feed NaOH to the wastewater in proportion to plant flow at a rapid mixer at the downstream end of the Parshall flume. The time of flow between the point of addition of $FeCl_2$ and NaOH is approximately 10 seconds.

The chemically dosed wastewater will be divided and will flow into two new air agitated mixing basins downstream from the Parshall flume. Each basin will be 17.5 ft-wide by 38.7 ft long and will have a sidewater depth of 12.6 ft. At the maximum plant flow rate of 18 mgd, approximately 10 minutes detention will be provided. At the average plant flow rate of 12 mgd and with one basin out of service, there will be approximately 7.5 minutes detention. Because some material may settle in these basins, screw conveyors will be installed in each to remove such material to a sump. Settled material will be pumped to the downstream side of these basins to be re-mixed with the plant flow. Stop plates and drains will allow each basin to be removed rapidly from service for maintenance.

The polymer facilities are designed for use of Dow Purifloc A23. Two 1,500 gal. solution tanks, each equipped with a mixer, will be included. Two gear-type pumps will be used to feed polymer in proportion to plant flow A third standby pump will also be provided. Polymer will be added at rapid mixers downstream from the air agitated basins. The wastewater will then flow into flocculation basins that are gently stirred by air for 1 to 5 minutes and then into settling basins. Three centrifugal blowers will be provided; one will supply air for agitation of the two mixing basins, one will supply air for agitation of the flocculation basins, and the third will be a standby. A small air line will be run to the ferrous chloride storage tanks to allow intermittent agitation. All chemical storage tanks, day tanks, and pump feeders are located in a heated building.

It is anticipated that sludge from the expanded plant can be handled by existing equipment. At present, sludge is stored in holding tanks until a sufficient amount has accumulated for operation of the filters and incinerator. The sludge is conditioned with polymer prior to filtration on two 150 ft^2 vacuum filters. Filter cake is then incinerated in a 14.25 ft diam., six hearth incinerator.

4.7.3 Wyoming, Michigan

An expansion of the Wyoming, Michigan wastewater treatment plant is being designed for 80% phosphorus removal and increased organic removal. The plant will include new chemical mixing and flocculation basins, two primary settling basins, a chemical feed and storage building, secondary aeration basins (trickling filters are to be used for roughing), final settling basins, and sludge handling facilities. Phosphorus will be removed by addition of ferric chloride to the new chemical mixing basins, followed by polymer addition, flocculation, and sedimentation in the expanded primary facilities. Provision will be made to feed ferric chloride and polymer to the final clarifiers for increased phosphorus or solids removal. All chemical tanks and feed equipment will also be designed to allow the use of liquid alum, ferrous chloride, ferrous sulfate, and sodium hydroxide.

Plant components associated with phosphorus removal are to be designed on the following basis. The chemical mix and flocculation basins will have a combined detention time of 20 minutes at the average design flow of 19 mgd. At design flow, the two existing and two new primary basins will have overflow rates of 604 gpd/ft^2, weir rates of 15,100 gpd/ft, and detention times of 2.07 hr.

Capacity of the sludge handling facilities will be more than doubled. The present units consist of two holding tanks with mixers, two 250 ft^2 vacuum filters, and one 6 hearth incinerator rated at 1,500 lb dry solids/hr. The expanded sludge processing facilities have been designed on the following basis:

Suspended Solids		
Plant load, lb/day	63,460	
Overall removal (95% assumed), lb/day		60,290
Biochemical Oxygen Demand		
Plant load, lb/day	48,260	
Removed by aeration basins, lb/day	7,850	
Waste activated sludge, lb/day*		3,925
Total		64,215

Solids added by lime and ferric chloride conditioning (15%)	9,635
Total solids loading, lb/day	73,890
Total solids loading, lb/hour	3,080

*Waste activated sludge production based on 0.5 lb solids generated per lb BOD removed.

Additional equipment which will be provided to handle this load includes two air flotation thickeners, two sludge holding tanks, two vacuum filters, and one incinerator. The two air flotation thickeners will each have an effective area of 1,170 ft^2. The sludge will be conditioned with lime or ferric chloride prior to flotation thickening at a rate of 20 $lb/ft^2/day$. The two holding tanks will allow 4 days storage capacity for the additional sludge. Each of the two new 500 ft^2 vacuum filters will have a capacity of 4.2 $lb/ft^2/hr$, and one will be on a standby basis. The six hearth incinerator will handle 2,100 lb of dry solids per hr. Lagoons will be provided for disposal of the ash.

4.8 Costs

All cost estimates for the removal of phosphorus by chemical precipitation-flocculation during primary treatment are dependent to some extent on raw wastewater flow rates. The cost per 1,000 gal. for chemicals may be less in larger installations because of large volume purchasing at lower unit prices. Total capital costs for tankage, chemical feed pumps, and instrumentation will be less for smaller plants. Placed on a unit volume of waste treated basis, however, capital costs for small plants will be higher than for large plants.

4.8.1 Chemical Costs

The chemical costs for phosphorus removal in the primary settler have been computed for both liquid alum and liquid $FeCl_3$. The respective unit costs of these chemicals were estimated at $60/ton for alum (on a 17.1% Al_2O_3 basis) and $90/ton, as $FeCl_3$, for liquid $FeCl_3$. These figures represent average delivered prices.

As a basis for estimating costs in terms of ¢/1,000 gal., an influent P concentration of 10 mg/l was assumed. For removal of 80% of the total phosphorus in primary treatment, equal molar ratios of Al:P and Fe:P of 1.5:1 were assumed. Polymer costs were included, based on a dosage rate of 0.5 mg/l and a chemical cost of $1.50/lb. The chemical costs of phosphorus removal treatment are estimated to be 3.6 ¢/1,000 gal. for liquid $FeCl_3$ and 4.2 ¢/1,000 gal. for liquid alum, respectively. As mentioned earlier, these costs may vary depending on the size of treatment plant and cost of freight.

The chemical cost for phosphorus removal at plants where waste pickle liquor is available is more difficult to predict accurately because of varying iron concentrations, larger storage requirements, freight charges, and the additional alkalinity requirement.

4.8.2 Capital Costs

Equipment costs for phosphorus removal with aluminum and iron are small relative to the chemical costs. Costs for facilities will be presented in Chapter 10 for single point addition of several different chemicals. For convenience, however, capital costs of a typical primary addition process using liquid ferric chloride and polymer are given below in Table 4-7.

Table 4-7

STORAGE AND FEEDING CAPITAL COSTS FOR LIQUID FERRIC CHLORIDE AND POLYMER ADDITION BEFORE THE PRIMARY

Plant Size	Capital Cost	
(mgd)	(Total $)	(¢/1,000 gal.)*
1	20,810	0.446
10	71,740	0.154
100	697,000	0.149

*Based on 25-year amortization at 6%.

The capital cost expressed on a ¢/1,000 gal. basis allows comparison with the chemical costs given earlier and illustrates that the capital costs are not very significant. Capital costs for alum addition are the same or less than those for liquid ferric chloride addition, depending on plant size. Further information on costs and on the design basis adopted for determining capital costs is given in Chapter 10.

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Chapter 5

PHOSPHORUS REMOVAL BY LIME ADDITION BEFORE THE PRIMARY SETTLER

5.1 Description of Process

Addition of lime to raw wastewater for the purpose of phosphorus precipitation can be applied to conventional plants using the activated sludge process. Because the pH of the primary effluent ahead of the activated sludge process should not exceed a value of 9.5 to 10.0, phosphorus removal by this method is generally limited to efficiencies of about 80%. Higher pH values can result in biological upset. Greater phosphorus removal, if necessary, may be accomplished by aluminum or iron addition to the aerator or final settling basin. One fortunate aspect of the lime clarification-activated sludge process is that the high pH of the primary effluent is reduced in the aeration tank by the CO_2 produced by biological metabolism.

The removal of phosphorus in primary treatment is accompanied by increased BOD and suspended solids removal and thereby reduces the organic load on the secondary treatment facilities. Reduced organic load benefits overloaded plants and also aids in establishing nitrification. Some phosphorus removal is accomplished by biological treatment and this method takes advantage of that characteristic to remove some of the remaining phosphorus that would be more difficult to remove by chemical treatment. Use of lime rather than an iron or aluminum salt does not add large amounts of other ions such as sulfate or chloride to the treated water. Other possible benefits include improved oil, grease, and scum removal and less corrosion in the primary sludge handling system.

One consideration in selecting the lime clarification-activated sludge process is the ability to use existing primary settling basins for phosphorus removal. The initial capital expenditure for phosphorus removal, therefore, may be comparatively small. There is a need for feeding lime, mixing of the lime with the water, and flocculation. Lime feeding equipment is discussed in Chapter 10. Equipment requirements for mixing and flocculation will vary depending on the configuration of the existing plant. In some cases, mixers and flocculators may have to be provided. In others, satisfactory flocculation will occur with no additional equipment.

Sludge disposal may become more complex and lime sludge may not be compatible with existing disposal processes such as anaerobic digestion. Furthermore, recovery of lime by recalcining which is discussed in Chapter 8 may not be practical with this process because of the low $CaCO_3$ content of the sludge.

5.2 Process Performance

The concept of using lime for phosphorus removal in primary treatment preceding activated sludge was advanced a number of years ago. Since then it has been investigated at laboratory scale and full plant scale.

Albertson and Sherwood (1) established with laboratory data that lime treatment of raw wastewater is an efficient means of removing 80% or more of the phosphorus and 60 to 70% of the BOD. Discharge of the pH 9.5 or above water to a complete-mix activated sludge system eliminated the need for pH correction because of CO_2 production by the biological activity in the aerator.

Schmid and McKinney (2) investigated lime precipitation of phosphate from wastewater on a laboratory scale. Their studies also included precipitation of various phosphate compounds from distilled water solutions by the addition of lime and consideration of the dewatering characteristics of the sludge formed by phosphate precipitation. These investigators concluded that both orthophosphate and polyphosphates, normally present in wastewater, can be precipitated readily with lime. They also noted that the complex phosphates present in wastewater seriously inhibit calcium carbonate precipitation. In view of this, they concluded that sludge recalcination for lime recovery would not be practicable. For the wastewater employed in the test work, the addition of 150 mg/l of lime (calcium hydroxide) resulted generally in a pH of 9.5, total phosphorus removal before biological treatment of 60%, and suspended solids removal of 90%. The authors concluded that sludge produced by primary treatment, where lime is employed, may be about twice that obtained normally. Overall sludge production, including waste activated, would be about 1.5 times normal. The process can be controlled by operating at a constant pH. Operation of a complete-mix activated sludge system is not hindered, provided proper control is exercised. Microbial CO_2 production in the activated sludge unit is sufficient to maintain the pH near neutral. A mixture of lime-precipitated primary sludge and waste activated sludge will dewater and filter well with small additions of anionic polymers.

Following extensive laboratory jar test and bench-scale studies, Black and Lewandowski (3) conducted a plant-scale investigation of lime treatment for phosphorus removal at the 180,000-gal. (imperial) per day Richmond Hill, Ontario activated sludge plant. Lime was added before the primary clarifier on a continuous basis for approximately 3 months. In this case, no special mixing or flocculating equipment was provided. The raw wastewater had an alkalinity of 381 mg/l as $CaCO_3$, a total hardness of 401 mg/l as $CaCO_3$, and an average total phosphorus content of 10.6 mg/l as P. Normal plant operation produced a total phosphorus removal of 39% with BOD and suspended solids removals in the primary unit of 21% and 37%, respectively. A lime dosage of 175 mg/l was sufficient to produce a pH of 9.3. During lime addition, BOD and suspended solids removals in the primary clarifier were increased to 72 and 78%, respectively, and total phosphorus removal averaged 83%. Phosphorus removal over the total plant was 92% resulting in a total phosphorus concentration in the secondary effluent of 0.9 mg/l. Throughout the study, $CaCO_3$ accumulated on the effluent weirs and outlet troughs of the primary clarifier but was easily removed with a water hose. There was no appreciable buildup on submerged structures within the primary unit. Lime treatment had no detrimental effects on the secondary process. The high pH of the primary effluent was promptly reduced by the CO_2 generated in the aeration tank. Sludge equilibrium was reached and maintained throughout the study. As a result of improved primary performance, the organic loading to the secondary plant was reduced, resulting in its behavior as an extended aeration process with no secondary sludge being wasted. If sludge wasting had been required this would have increased the phosphorus removal.

The sludge precipitated in the primary clarifier was a mixture of a calcium phosphate compound, $Ca(OH)_2$, $CaCO_3$, and coagulated organics plus grit. This sludge was free flowing and considerably different from the gelatinous type sludge produced in bench-scale studies. The sludge concentrated readily to 10 to 12% solids by gravity thickening but was very difficult to dewater. The sludge removed from the primary unit had a solids content of 6 to 8%. The investigators concluded that there would be no practical value in attempting to digest the sludge derived from the lime treatment process, inasmuch as biological activity of the sludge is suspended by the high lime content. The sludge is free from noxious odors even after prolonged holding under anaerobic conditions. The authors suggest that the sludge might be dewatered by polymer conditioning and vacuum filtration or centrifugation followed by landfill disposal.

The study results indicated that the organic capacity of an existing plant could be increased without jeopardizing good treatment through the addition of lime to the primary units. Laboratory studies using a combined flocculation-sedimentation system with high rates of primary sludge return showed that recirculation of the lime sludge increased the efficiency of the lime and improved effluent clarification. Supernatant from the chemical sludge, even after 5 weeks of storage, contained no appreciable amount of phosphorus, either total or soluble.

It was concluded that the lime treatment process could be readily controlled in response to flow or on the basis of pH in a flash mixer or in the primary effluent. It was also concluded that lime control in this manner would be simpler than that of a chemical dependent on incoming phosphorus concentration.

5.3 Example of a Plant Design

During the period of January to April 1967, treatability studies for phosphorus removal were made on the influent to the Rochester, New York treatment plant (4). At this time the raw wastewater phosphorus concentration averaged about 8 mg/l as P. Studies showed that the activated sludge process removed about 20% of the influent phosphorus. This removal agreed reasonably well with the phosphorus uptake normally experienced with biological metabolism.

Jar tests were performed to determine the chemical treatment necessary to achieve the State requirements of 80% phosphorus removal. Lime, and ferric chloride with anionic polymers were used in the jar tests. Either 100 mg/l of lime as CaO (pH 9.5) or the combination of 40 mg/l $FeCl_3$ and 0.5 mg/l Dow A23 was found to reduce the phosphorus concentration of the supernatant by about 70%. Adding the removal from the biological treatment would give an overall phosphorus removal of at least 80%. BOD and suspended solids reductions indicated by these jar tests were 50% and 80 to 90%, respectively. Lime was selected as the chemical to be used.

The present treatment plant facilities consist of settling tanks and Imhoff tanks, sludge holding tanks, coil filters, and sludge incinerators.

Expansion plans call for modification of present settling basins with flocculation sections, addition of settling basins incorporating flocculation sections, conversion of the Imhoff tanks to complete-mix activated sludge units, chlorination, and discharge to Lake Ontario through a new 10 ft diameter 18,000 ft outfall line. Detention time in the flocculation sections of existing settlers will be 20 minutes. In the new basins, the flocculation time will be 40 minutes. Existing and new settlers will have overflow rates of 1,000 gpd/ft^2 and detention times of 2.3 and 2 hours, respectively, at the average design flow. A schematic diagram of the expanded plant is shown on Figure 5-1. Plant design is based on the following criteria:

Average design flow, mgd	100
Maximum flow, mgd	200
Minimum flow, mgd	50
BOD loading, lb/day	250,000
BOD concentration, mg/l	300
Suspended solids concentration, mg/l	300
Volatile suspended solids concentration, mg/l	240

Lime will be added just after the waste passes through the comminutors. The pH will be monitored ahead of the flocculation basins and lime addition will be automatically regulated to maintain a pH of 9.5.

Anionic polymer will be fed at a point just ahead of the flocculation basins when necessary to improve settling. Provisions for returning solids from the primary basins to the flocculation basins have been included to enhance the flocculation process.

Although lime treatment together with biological treatment should remove 80% of the incoming phosphorus as required by the State, chemical facilities for addition of alum to the aeration tanks will also be provided. This system of phosphorus removal may be used alone when the lime system is not in operation or may be used to supplement lime clarification for greater phosphorus removal.

Since 80 to 90% suspended solids removal is expected in the primary basins, the quantity of primary sludge based only on the increased suspended solids removal will be about 1-1/2 to 2 times the normal sludge quantities. The sludge produced as a result of chemical precipitation must be added to this quantity. It is anticipated that, including the chemical sludge, about three times the normal amount of sludge will be produced with chemical treatment. The primary sludge quantities at design flow with lime treatment are estimated to be 175 tons/day. Without lime treatment the primary sludge quantities will be about 65 tons/day. Since the load on the activated sludge portion of the process will be reduced by about 50%, the quantity of waste activated sludge, which is very difficult to dewater, will also be reduced proportionally. At design loads with lime treatment, the waste activated sludge quantities are estimated to be about 30 tons/day as compared to about 75 tons/day if lime treatment were not used. The solids handling facilities have been designed for a capacity of 270 tons/day.

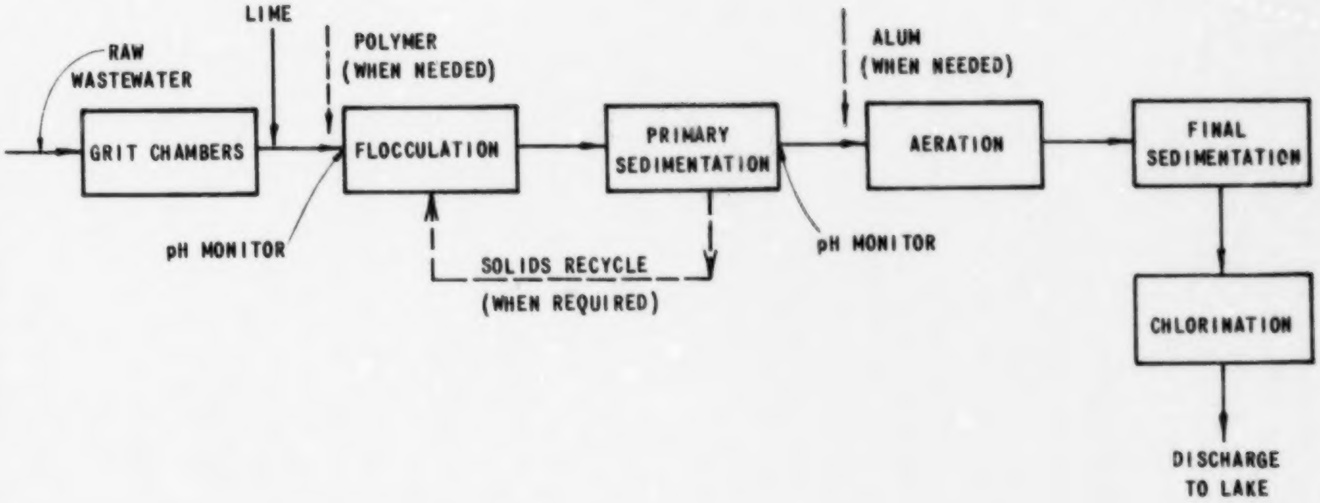
RAW WASTEWATER
LIME
POLYMER (WHEN NEEDED)
GRIT CHAMBERS
FLOCCULATION
PRIMARY SEDIMENTATION
ALUM (WHEN NEEDED)
AERATION
FINAL SEDIMENTATION
pH MONITOR
SOLIDS RECYCLE (WHEN REQUIRED)
pH MONITOR
CHLORINATION
DISCHARGE TO LAKE

FIGURE 5-1 SCHEMATIC OF PROPOSED PHOSPHORUS REMOVAL PROCESS AT ROCHESTER, NEW YORK

The sludges will be combined and thickened in gravity thickeners to a minimum concentration of 5%. The thickeners were designed for a solids loading of 13 lb/day/ft^2. After thickening, the sludge will be dewatered on vacuum filters and incinerated by multiple hearth furnaces. Incinerator ash, which will contain some CaO and other calcium and phosphorus compounds, will be sluiced and pumped to two, 0.5 acre lagoons, each having a water depth of 11 ft and operating in series. The phosphorus is not expected to leach out because of the high pH in the lagoons. The effluent from the lagoons will be pumped back to the head end of the plant. It is expected that the lagoons will have to be emptied every 9 to 12 months.

5.4 Capital and Operating Costs

The cost of adding rapid mix and flocculation facilities to existing plants has been estimated for 1, 10, and 100 mgd plants. These estimates should serve as a general guideline where phosphorus is to be removed by lime addition to the primary settler.

Estimated costs for rapid mix facilities include new concrete basins, having a detention time of 1 minute at the design flow, and new mixers, with shafts and impellers constructed of stainless steel. Flocculation facilities consist of the addition of the necessary baffle walls and mixers to provide a flocculation chamber, having a detention time of 30 minutes, within existing circular settling basins. Estimated costs of these additions are given below and include an allowance for the Contractor's installation, overhead, and profit, plus an additional allowance of 20% of the construction cost for engineering and contingencies.

Unit	Plant Size, mgd 1	 10	 100
		(Estimated Costs, $)	
Rapid Mix	1,900	11,300	71,100
Flocculation	55,700	229,700	855,200
Total	57,600	241,000	926,300

The cost per thousand gallons for adding single-stage lime treatment to an existing plant was determined by amortizing the capital cost at 6% over 25 years and adding the cost of chemicals. Operational expenses such as labor and power are not included. Capital investment for adding rapid mixing and flocculation equipment are given above. Investment figures for the lime storage and feed systems are given in Chapter 10, Section 10.3.3. Lime costs per 1,000 gal. are based on a chemical cost of $20/ton with 90% available CaO at a dosage of 150 mg/l as CaO. The equipment and chemical costs (¢/1,000 gal.) are given below:

Item	Plant Size, mgd		
	1	10	100
	(Cost, ¢/1,000 gal.)		
Equipment Amortization			
Rapid mix and flocculator	1.2	0.5	0.2
Chemical storage and feed system	0.4	0.2	0.1
Chemical Cost	1.4	1.4	1.4
Total Cost	3.0	2.1	1.7

Costs listed in the table above do not include an allowance for additional equipment for sludge handling and disposal and corresponding operating costs. Since a considerable quantity of additional sludge is produced, this expense is likely to be substantial.

5.5 References – Chapter 5

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Chapter 6

PHOSPHORUS REMOVAL IN TRICKLING FILTERS BY MINERAL ADDITION

A logical approach to removal of phosphorus in trickling filter plants would be the use of precipitating chemicals somewhere in the system. Iron and aluminum can both be used for this application. Because of their similar action, iron and aluminum salts are commonly referred to collectively as minerals in wastewater treatment and are considered separately from calcium (lime). Required physical facilities for adding iron and aluminum are relatively simple. They are covered, along with special considerations for specific chemicals, elsewhere in this manual. This discussion will emphasize the merits of alternative addition points, review methods for determining dosage, and report studies where chemicals have been dosed directly to a trickling filter and to trickling filter effluent. Broader aspects of mineral addition for phosphorus removal are covered in other sections of this manual and in several recent reports (1, 2, 3, 4).

6.1 Pre-Design Decisions

State quality standards on phosphorus usually form the major basis for considering mineral addition for a given trickling filter plant. However, phosphorus concentration in the effluent is only one factor involved; effluent BOD and suspended solids will also be markedly reduced by this form of chemical treatment.

Modification of existing trickling filter plants for phosphorus removal is often a simple operation, and inclusion of chemical treatment facilities in new plants is a relatively minor addition. In either case, capital costs for chemical treatment equipment are relatively low. Also, the required degree of phosphorus removal has little effect on capital cost. Facilities for 80% phosphorus removal are identical to those for 95% removal.

On the other hand, operating costs are substantially higher to obtain successful treatment with chemical addition. More constant surveillance and more laboratory support are needed than with a conventional plant. Added operating costs are on the order of 4¢ to 5¢ per 1,000 gal. for reducing phosphorus (as P) to 2 mg/1 with an accompanying reduction of BOD and solids to 15 mg/1. For an additional 1¢ to 2¢ per 1,000 gal., phosphorus can be reduced to 1 mg/1 or less and BOD and solids to 10 or 12 mg/1. This simultaneous reduction of phosphorus, suspended solids and BOD may be the deciding factor for adopting chemical treatment in a trickling filter plant.

6.2 Process Options

There are three locations within a trickling filter plant where chemical precipitation can be incorporated: the primary clarifier, the trickling filter, and the final clarifier. The preferred approach is to provide facilities to serve both the first and last of these but not to add chemicals directly to the trickling filter.

Chemical addition to primary clarifiers will produce the greatest sludge yield of any one arrangement. Since treatment at this point will reduce influent BOD about 50%, it may be an appealing approach in an existing plant which is organically overloaded. Unfortunately, a significant amount of the phosphorus present may not be in the ortho form at this stage and may not be effectively precipitated and removed. Even after chemical treatment, primary effluent may remain fairly turbid, but will clear up during passage through the rest of the plant.

Mineral addition in the trickling filter does not seem to cause any functional problems although the filter may blotch and slough more than usual. A modest inorganic film often forms over the surface of the rocks but does not seem to interfere with biological activity or cause ponding. As data shown later will indicate, however, high degrees of phosphorus removal are not attained when chemicals are added only to the filter. Poor removal may result partly from short circuiting in the filter. This approach is not recommended.

Treatment in the final clarifier alone is a risky approach. If the operation is not done effectively, poorly treated effluent can escape to the receiving water. In spite of these reservations, experience has shown that precipitation in the final clarifier can be both effective and controllable. Orthophosphate predominates at this point and it precipitates readily. Also, bio-degradable detergents which can interfere with precipitation are largely absent. If underflow solids from a dosed final clarifier are returned to the primary clarifier, it will stimulate unusually effective clarification there.

In order to allow for incorporating the advantages of chemical treatment in both the primary and secondary clarifiers, it is recommended that chemical addition and mixing facilities be provided at both locations. Facilities required are not expensive and provision for mineral addition to both primary and secondary clarifier gives considerable flexibility of operation. Chemical treatment in both clarifiers can then be used if experience shows it to be the most effective technique at the particular plant involved.

6.3 Performance Data Using Aluminum Salts

Both alum and sodium aluminate have been tested for phosphorus removal in trickling filter systems. Where alkalinity is low, alum may depress the pH too far to give good results. In these cases sodium aluminate would be preferred over alum.

The Fairborn, Ohio Wastewater Treatment Plant was used in a study (5) of direct dosing to a trickling filter with sodium aluminate. However, the maximum phosphorus removal obtained was only 66%. This emphasizes the need for addition at other points of the process in order to obtain high reductions in phosphorus. Data on phosphorus removals obtained, with both a dosed and control trickling filter, are shown in Table 6-1.

Table 6-1

FAIRBORN, OHIO – TRICKLING FILTER PHOSPHORUS REMOVAL

Al:P weight ratio of 1:1 in dosed filter

Process Stream	Total Phosphorus (mg/l)		% P Removed
	Range	Average	
Raw wastewater	9.2-18.0	13.2	–
Primary effluent	9.8-17.0	12.5	5.3
Final effluent, dosed	2.8- 6.2	4.5	66.0
Final effluent, control	9.0-12.8	10.9	17.4

Al:P weight ratio of 1.6:1 in dosed filter

Process Stream	Total Phosphorus (mg/l)		% P Removed
	Range	Average	
Raw wastewater	10.2-36.0	14.8	–
Primary effluent	7.7-12.8	11.3	23.6
Final effluent, dosed	4.6- 8.4	5.2	64.8
Final effluent, control	8.0-14.2	10.6	28.4

Sludge analyses from the sodium aluminate studies revealed a volatile content of about 70% for the control filter sludge and 50% for the dosed filter sludge. The combined sludge from one dosed and two normally operating filters was digested anaerobically. As a result of only one of the three trickling filters being dosed, the excessive inert sludge bulk reported by Dean (6) did not occur in the digesters.

The addition of alum to trickling filter effluent before final settling is being studied at Richardson, Texas (7) (addition of iron and polymer at Richardson is covered later). A plant schematic is shown on Figure 6-1 along with the points of chemical addition. At design capacity of 1.6 mgd the treatment units have the following design characteristics:

Primary Clarifiers	
Detention time, hours	3.9
Overflow rate, gpd/ft^2	424
Trickling Filters*	
Hydraulic loading, gpd/ft^2	95
Process loading**, lb BOD_5/1,000 ft^3	11.9
Depth, ft	6.5
Recirculation, gpd	70,000
Final Clarifier	
Detention time, hours	2.6
Overflow rate, gpd/ft^2	416

* average

** assuming 30% BOD removal in primary

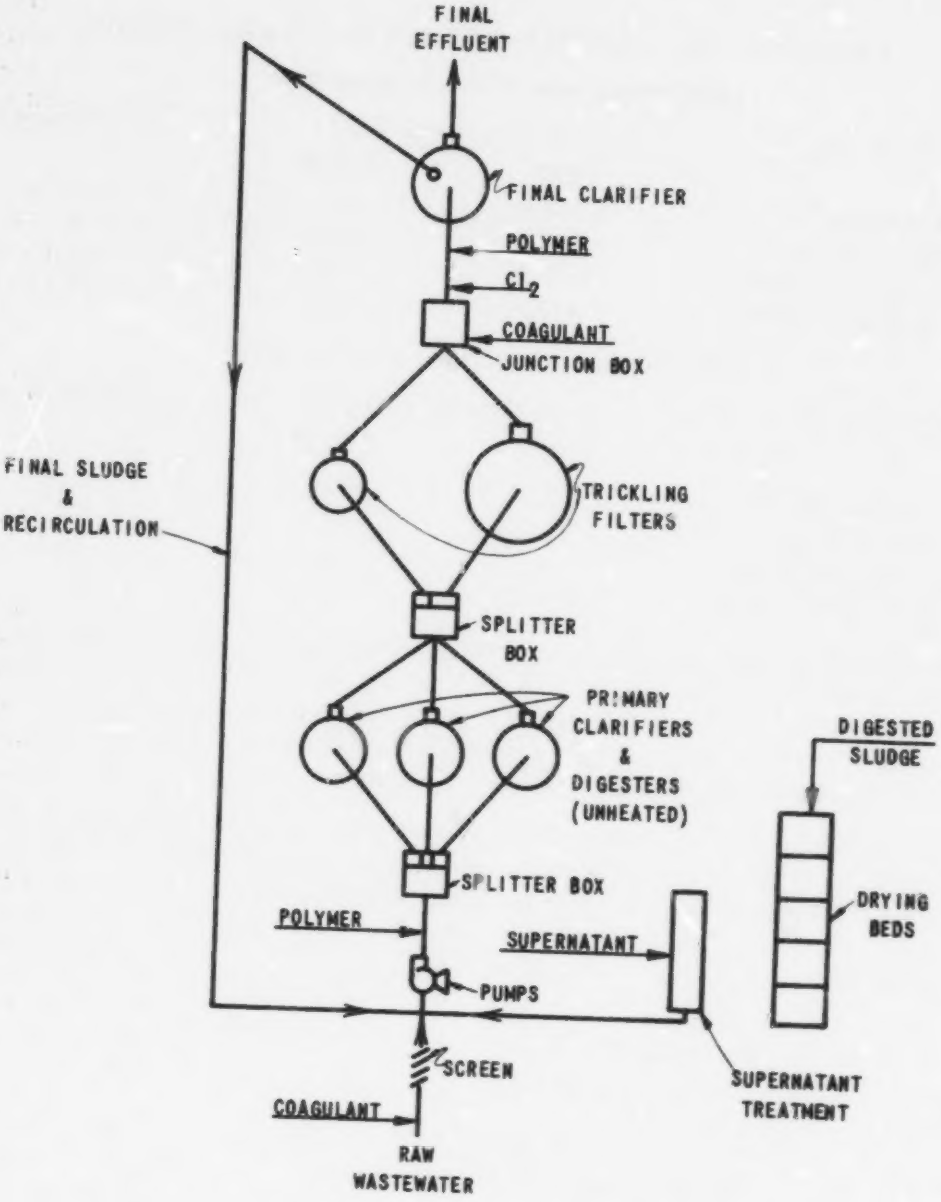
FINAL EFFLUENT
FINAL CLARIFIER
POLYMER
Cl_2
COAGULANT
JUNCTION BOX
FINAL SLUDGE & RECIRCULATION
TRICKLING FILTERS
SPLITTER BOX
PRIMARY CLARIFIERS & DIGESTERS (UNHEATED)
DIGESTED SLUDGE
SPLITTER BOX
POLYMER
SUPERNATANT
DRYING BEDS
PUMPS
SCREEN
SUPERNATANT TREATMENT
COAGULANT
RAW WASTEWATER

FIGURE 6-1 RICHARDSON, TEXAS TREATMENT PLANT FACILITIES

Sludge is digested in the lower compartment of each primary clarifier prior to dewatering on sand beds. Underdrain filtrate from the beds and digester supernatant, after treatment with alum, are returned to the head of the plant. Major modifications to the plant included chemical storage and handling facilities and a flash mix unit for the junction box after the trickling filters. Chemicals are also added and mixed in the raw wastewater wetwell.

Since the study will not be completed until early in 1972, the data and inferences presented here are preliminary and not the final results. Influent phosphorus has been about 8 mg/l (110 lb/day) of which about 70% is soluble. An Al:P weight ratio of 1.83:1 (molar ratio 2.1:1), fed exclusively after the trickling filters, reduced the effluent phosphorus concentration to 0.5 mg/l.

A split feed of 20% to the raw wastewater and 80% after the trickling filter at an Al:P ratio of 1.74:1 (molar ratio 2:1) reduced effluent phosphorus to 0.3 mg/l.

6.4 Performance Data Using Iron Salts

Addition of iron to trickling filters has been studied at Detroit and Wyoming, Michigan and at Richardson, Texas.

The Detroit Metro Water Department (8) compared the performance of pilot activated sludge and plastic-media trickling filter systems in order to decide on the design basis for a full-scale plant. Chemicals were added before the primary settler for both the trickling filter and activated sludge studies, and to the primary effluent for the activated sludge study. The plastic-media trickling filter test program consisted of a series of 16 experimental runs. Application rates ranged from 1.0 to about 3.5 gpm/ft^2. The raw wastewater inorganic phosphate averaged 7.9 mg/l P with a range of 1.9 mg/l to 26.6 mg/l. Without the addition of chemicals, total phosphate removals ranged from 18 to 57%. With the addition of 15 mg/l of iron (pickle liquor) removals were 34 to 77%. With 15 mg/l iron and 0.3 mg/l Dow Purifloc A23 polymer, removals ranged from 47 to 80%. With 15 mg/l iron, 0.3 mg/l A23 polymer, and 30 mg/l NaOH, removals were in the range of 67 to 78%. It was concluded that control of conditions for removal of phosphorus using the plastic-media trickling filters would be more difficult than in the activated sludge process. The activated sludge process achieved more consistent and higher phosphorus removals with iron addition and was recommended for the full-scale plant.

Dow Chemical Company performed plant scale iron addition tests at the Wyoming, Michigan trickling filter plant (9, 10, 11). The 1969 study was designed for ferric chloride addition to the raw wastewater to take advantage of the additional BOD and suspended solids removal in the primary. However, due to the recirculation of the trickling filter sludge with its high phosphorus content, it was necessary to split the $FeCl_3$ addition between the raw wastewater and the trickling filter effluent. The chemical feed and phosphorus removal data are presented in Table 6-2.

Table 6-2

IRON ADDITION FOR PHOSPHORUS REMOVAL AT WYOMING, MICHIGAN

	Test No.			
	1	2	3	4
Chemical Feed				
Fe^{3+} to raw wastewater, mg/l	15	15	15-20	15
Fe^{3+} to trickling filter effluent, mg/l	5-7	3-4	0	7
Purifloc A23, mg/l, before primary sedimentation	0.4	0.3	0.2-0.3	0
Total Phosphorus				
Raw, mg/l	10.4	11.1	14.0	17.5
Primary effluent, mg/l	1.68	2.10	3.4	6.5
Removal, %	83.9	81.0	75.7	62.9
Final effluent, mg/l	0.88	1.46	2.9	5.7
Removal, %	91.7	87.6	79.2	67.4

To achieve the 80% phosphorus removal required by the State, iron addition of 15 mg/l to the raw wastewater and 3 to 4 mg/l Fe^{3+} to the trickling filter effluent along with 0.3 mg/l Purifloc A23 were necessary. Removal of BOD during chemical treatment ranged from 60% to 87% across the plant. Suspended solids removal in the primary averaged 75% with an overall removal of 87%. Iron addition to the Wyoming trickling filter plant permitted greater than 80% phosphorus removal only with the split chemical addition.

Initial data on iron addition at Richardson, Texas are included as examples in following discussions on trickling filter plant dosage and iron leakage.

6.5 Choice of Chemical Addition

Since in-plant chemical precipitation is a new technique, pilot-scale data should be obtained prior to design of new facilities. However, when the plant already exists, full-scale trials may not be any more expensive than pilot-scale tests and they will be more reliable.

The jar test is a most important tool for both designer and operator when using chemical addition. Information can be obtained inexpensively and in a short period of time, however, certain precautions should be taken. An auxiliary flash mix is highly desirable if the jar test apparatus does not have capability for true high energy mixing. After intense dispersion of trial chemicals, flocculation can be simulated by careful programming of time and mixing energy levels so they approximate conditions in the plant. The most effective way to match jar test and plant conditions is to adjust the jar test mechanism until it appears to duplicate the hydraulic regime observed directly in the plant. Duration of each mixing interval should approximate plant conditions, assuming plug flow. Finally, the machine should be left turning very slowly to approximate settling conditions in the plant. A slight motion in the test jars is a better representation of clarifier conditions than if stirring paddles are turned off. No matter how carefully the rotation speed and timing intervals are selected to approximate plant conditions, the operator should devote considerable practice to use of the apparatus before his results are accepted and used.

A wide variety of other bench scale and laboratory tests (12, 13) hold considerable promise in wastewater technology. At this time, however, none of them have proved as effective as the jar test.

6.6 Nature and Role of Chemicals Involved

Compounds mentioned here are those most likely to be selected for chemical treatment in a trickling filter plant. Of the commercially available metal salts (14), liquid forms are more convenient to handle, especially at small plants. They may have the lowest overall cost and are generally more effective than dry forms. However, it may be necessary to use dry salts because transportation of pre-mixed solutions, containing both water and salt, may be costly.

Metal salt forms include liquid ferric chloride, pickle liquor, liquid alum, and sodium aluminate. Information on properties and handling these chemicals is given in Chapter 10.

Polymers are available in a variety of forms, mostly dry powders. Since the amount of polymer required is far less than the amount of metal salt, the problems of using dry polymer compounds are not severe. Of the three categories of polymers (cationic, anionic and nonionic) there is no universal choice regarding the most effective type.

Whenever chemicals are chosen, they must perform two main functions: precipitate phosphorus, and remove all types of colloids in the water. Phosphorus precipitation is caused solely by metal salts and involves conversion of soluble orthophosphate to insoluble phosphate colloids. Following that, both metal salts and polymers serve to coagulate (destabilize) colloids. Flocculation of destabilized colloids is also brought about by both metal and polymers.

6.7 Dosage Selection and Control

Proper dosage selection and control are major keys to the success of chemical treatment because this is where both cost and performance are determined. The technique selected must be effective but also should be simple and easily interpreted by the operator.

In the metal salt system, the key parameters are the mole ratio fed and the effluent phosphorus concentration. As a basis for dosage determination, complete composite hourly and daily phosphorus analyses should be performed initially. These data will provide a guide to both total daily dose required and to daily variations in dosing rate.

Once accomplished, this task need not be repeated because subsequent refinements can be based on phosphorus concentration in the final effluent. Phosphorus observations should be made at both the primary clarifier and the final sedimentation tank if direct dosing of the filter is discounted.

A suggested procedure for setting the dose is as follows. First, express incoming phosphorus in pounds per day, then convert to lb-mole/day by dividing by the atomic weight (lb/lb-mole). Atomic weights are 31 for phosphorus, 27 for aluminum and 56 for iron. Next set the mole ratio (metal:P) at the proper value, between 1.5:1 and 2:1, and compute the daily coagulant dose. An example calculation is shown below:

If incoming phosphorus (as P) is 310 lb/day,

$$310/31 = 10 \text{ lb-moles/day}$$

If desired mole ratio (M:P) is 2:1,

$$\left(\frac{2}{1}\right)(10) = 20 \text{ lb-moles metal required}$$

Using liquid alum having 4.37% Al,

$$(20)(27) = 540 \text{ lb Al required}$$

$$(540)/(11.1 \text{ lb/gal.})(0.0437)$$

$$= 1{,}100 \text{ gal. liquid alum required/day}$$

Then take the total volume to be fed per day and convert it into dosage rates with three to five changes of rate per day. These different feed rates should be selected to match incoming phosphorus at the point of injection. Dose must be matched carefully to phosphorus demand. Feeding twice the amount required part of the time and half the amount the remainder of the time will waste chemicals and lead to poor performance. In fact, overfeeding chemicals can give as poor results as underfeeding; this stems from the tendency of high metal concentrations to drive phosphorus colloids into a highly stabilized state. Dilution of sewage during rainy weather makes little difference in the amount of metal salts required since total phosphorus, in terms of lb/day, remains about the same.

Changes in feed pump settings can be made manually according to a schedule established and posted beforetime. Several types of cam regulated feed controls are also available and are effective (15).

After establishing the initial feed program, effluent phosphorus should be monitored hourly or continuously. The plant should be kept on a given feed schedule for at least three to five days, to allow the biological reactions to adjust to any effects of chemical dosing. The effluent phosphorus data will show peak concentrations resulting from dosing improperly for a period of time. The feed schedule can then be adjusted to remove these peaks. Finally, studies of different mole ratios fed can be made and the results plotted as shown on Figure 6-2. These curves can then be used to choose the mole ratio required to meet effluent phosphorus standards.

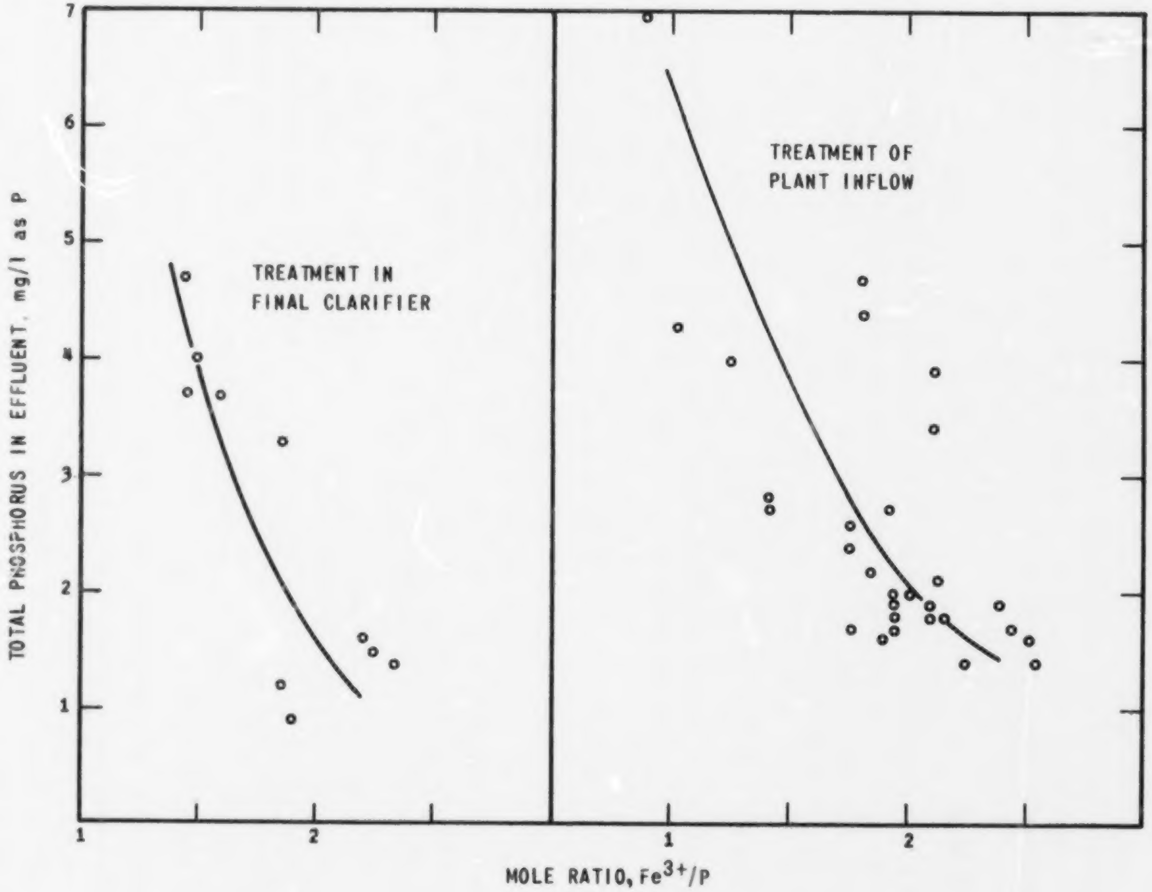

FIGURE 6-2 PHOSPHORUS REMOVAL BY IRON ADDITION (RICHARDSON, TEXAS)

Continuing operation of the plant should be monitored carefully and adjustments should be made in the chemical feed schedule when indicated by consistently high effluent phosphorus. Chemical feeding can be automated.

Polymer may be needed for good solids removal. Control of polymer feeding is relatively straightforward. Polymer feed rates are usually kept constant, but a reduced feed rate may be used during periods of low flow. Use of polymers may reduce metal salt demand and make metal addition more attractive. Polymers which prove effective in jar tests should be carefully evaluated in plant scale operation. Typical dosage is 1 mg/l or less. One direct way to judge polymer effectiveness is to observe effluent turbidity. Polymer feeding can also be automated.

6.8 Iron Leakage

Iron leakage may be a problem in trickling filter plants using iron precipitation. Figure 6-3 shows some typical data from Richardson, Texas (7). The problem appeared worse when iron was being fed in the final clarifier. It is evident that polymer treatment will reduce the escape of iron colloids.

6.9 Sampling and Analysis

The use of chemicals to remove phosphorus in trickling filter plants will require greater laboratory space, facilities, and reagents than normal treatment. Table 6-3 shows a typical matrix of data which is desirable at a conventional plant, as well as typical analyses required to support chemical treatment.

In the latter matrix, notice that the analyses include both anion and cation of the chemical being fed. Monitoring of alkalinity is optional unless the effluent concentration drops below 50 mg/l as $CaCO_3$. Some alkalinity is required to provide buffering and prevent a lowering of pH which would have an adverse effect on both chemical precipitation and biological activity.

Observation of turbidity in the final effluent is a good test. There is strong evidence that phosphorus and turbidity correlate in final effluents and the turbidity test provides a simple and responsive on-site control tool. Turbidity may be measured in the laboratory with a bench top meter or in the plant with a submersible disk lowered in the final clarifier. Observation of the height of the floc blanket in the clarifier is also a good means of control.

Effluent phosphorus concentration should be determined either on a semicontinuous basis with an automatic analyzer or on an hourly basis using automatic or manual sampling. In addition, a daily composite is also advisable. Sample portions should be flow-weighted rather than time-weighted. In plants enjoying stable effective performance, the schedule may be relaxed.

Signals from direct reading sensors and automatic analyzers can be used as a basis for automatic control of plant operation.

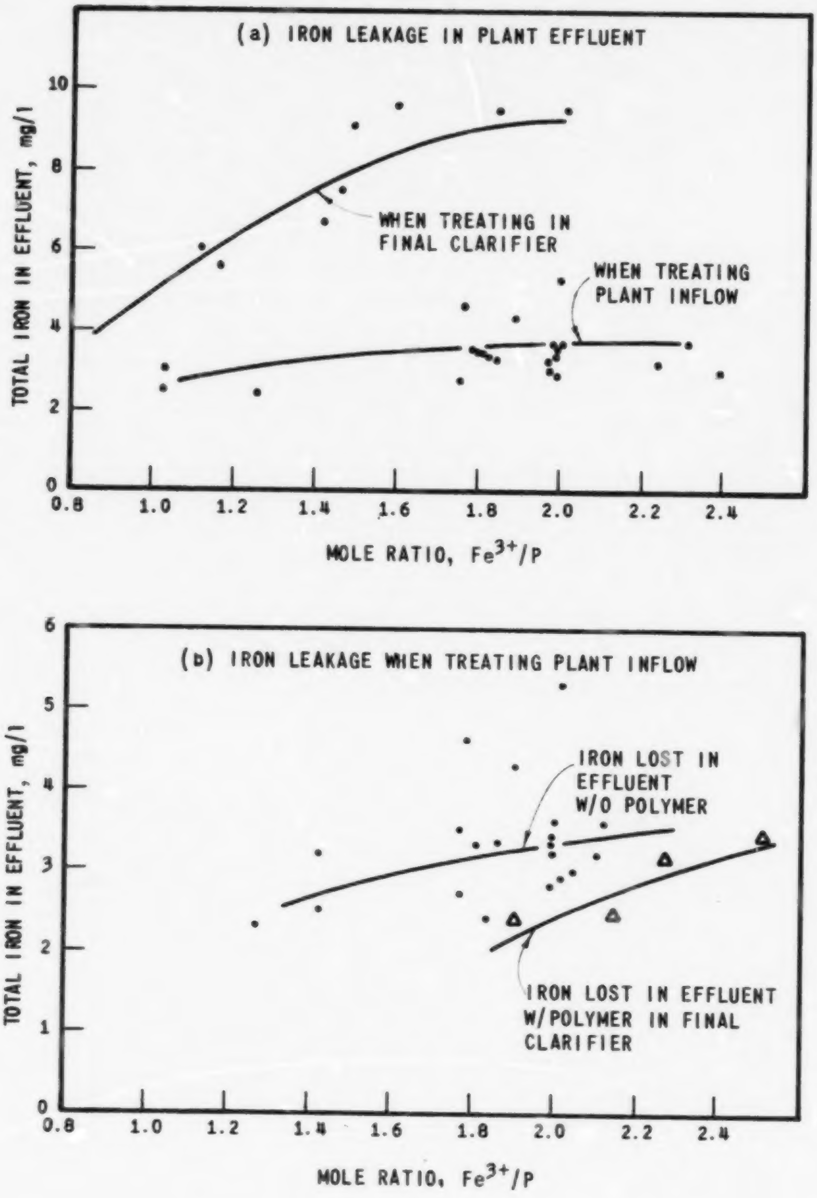

FIGURE 6-3 IRON LEAKAGE DURING PHOSPHORUS REMOVAL

ANALYSES FOR CONVENTIONAL TREATMENT

	RAW	PRIM EFF	FILT EFF	FINAL EFF	RECIRC	SLUDGE	SUPERNAT
FLOW	X				X	X	X
TOT SOL	X	X	X	X	X	X	X
TOT VOL SOL	X	X	X	X	X	X	X
SUS SOL	X	X	X	X	X		X
SUS VOL SOL	X	X	X	X	X		X
SET SOL	X	X	X	X			
BOD	X	X		X			X
DO	X			X			
COD	X	X	X	X	X		X
pH	X	X	X	X		X	X
TEMP	X	X	X	X		X	X

ADDITIONAL ANALYSES FOR CHEMICAL TREATMENT

	RAW	PRIM EFF	FILT EFF	FINAL EFF	RECIRC	SLUDGE	SUPERNAT
PHOS	X	X		X	X	X	X
ALK	X	X		X		X	X
FE	X	X		X	X	X	X
AL	X	X		X	X	X	X
SO_4	X	X	X	X		X	X
CL	X	X		X			
TURB	X	X		X			

TABLE 6-3 DESIRABLE LABORATORY ANALYSES

6.10 References – Chapter 6

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Chapter 7

PHOSPHORUS REMOVAL IN ACTIVATED SLUDGE PLANTS BY MINERAL ADDITION

7.1 Description of Process

The mechanism of phosphorus removal by mineral salt addition, usually iron or aluminum compounds, in a biological system is through a combination of precipitation, adsorption, exchange, and agglomeration as influenced by the pH and ionic composition of the water. The phosphorus removal technique is operationally simple and is accomplished by direct mineral addition to an aeration tank. Treatment costs are largely a function of the required effluent phosphate residual. Through optimization techniques, any degree of phosphorus removal may be provided. The main liability is the introduction of dissolved solids.

Phosphorus removal by biological systems alone, regardless of the internal mechanisms of phosphorus incorporation into the sludge solids, is limited, and removal is primarily controlled by the magnitude of solids wasting. The net amount of solids wasting is dictated by the quantity and type of substrate received and removed, system design and operation, and solids capture in the final effluent. A high rate system, with its greater solids production, would undoubtedly have the greatest background phosphorus removal potential by pure synthesis mechanisms. The magnitude of this background removal can be found by plant measurement with existing facilities or estimated by knowledge of the net solids production and phosphorus concentration in the waste activated sludge. Figure 7-1 shows the phosphorus removal capability for a variety of activated sludge systems at typically reported values for domestic wastewater treatment.

Historically, the benefits derived from mineral additions to activated sludge plants have been known since 1938 (1). Thomas (2) was among the first to point out the increased phosphorus removals. Wirts (3) also observed the increased phosphorus removals derived from a metal bearing industrial waste at a municipal treatment facility in Cleveland, Ohio. Early investigators of mineral addition for upgrading and modifying activated sludge systems for phosphorus removal include Tenny and Stumm (4), Barth and Ettinger (5), and Eberhardt and Nesbitt (6). Barth, et al., conducted pilot plant investigations with a three sludge system for combined chemical-biological nitrogen and phosphorus removal (7). Large scale, long term investigations have been recently completed at Manassas, Virginia (8), Pennsylvania State University (9), and Texas City, Texas (10). Recent publications have examined the use of aluminum (11) and iron (12) salts and investigated the kinetics and mechanisms of phosphorus removal using these minerals (13).

In the treatment of domestic wastewaters several investigators, most recently Rickert and Hunter (14), have shown that in a single sludge system a stable organic residual is found after aeration times of 30 to 60 minutes. Unfortunately, if treatment is stopped

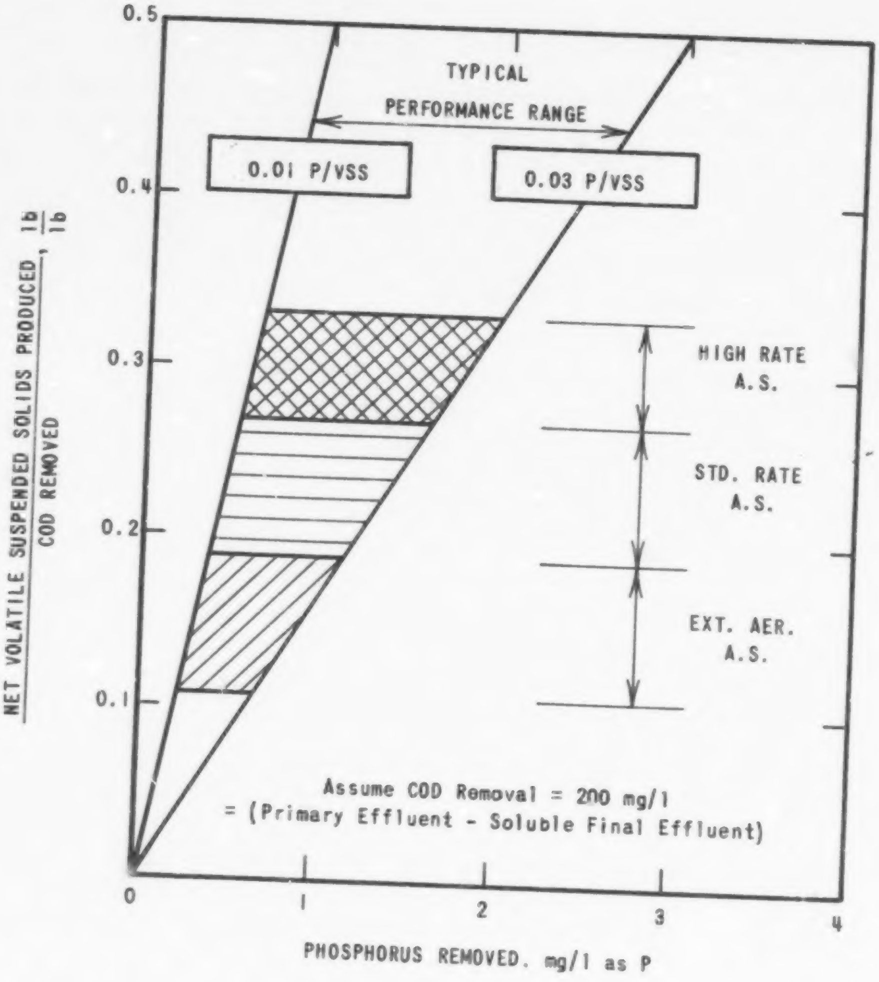

FIGURE 7-1 PHOSPHORUS REMOVAL CAPABILITY OF ACTIVATED SLUDGE SYSTEMS RECEIVING SETTLED DOMESTIC WASTEWATER

at this point, bio-flocculation is almost nonexistent and poor solids densities, settling characteristics and effluent quality result. However, when incorporating mineral addition, these liabilities are overcome and there exists an opportunity to derive capital savings from reduced tankage requirements to compensate for the increased operating expenditures associated with mineral addition. Indeed, in many cases, the use of a short residence time activated sludge system with mineral addition, settling, and filtration may result in a higher quality effluent for equal or lower costs than a purely physical-chemical approach of coagulation, settling, filtration and carbon adsorption (15). From a liability standpoint, however, the designer should be cognizant of the greater mass of generated waste activated sludge which must be successfully handled and treated for ultimate disposal.

Mineral addition to activated sludge for phosphorus removal offers the following advantages:

a. Ease of operation (direct chemical addition to existing unit processes).

b. Relatively small additional solids production (causing increases in sludge density and dewaterability).

c. Flexibility to changing conditions (treatment costs are largely a function of the influent and required final effluent phosphorus concentrations).

The main disadvantage is the introduction of dissolved solids which in some circumstances can be considered pollutants in their own right.

7.2 Mineral Selection and Addition

The selection of the mineral to use for phosphorus removal involves a multitude of considerations. Any commercial chemical, or industrial waste stream (for example, pickle liquor or alum sludge from a water treatment plant) bearing available aluminum or iron has potential application. Ideally, the best approach requires comparative full-scale plant tests of several months duration. This, of course, is usually impossible, and the next best approach is to run comparative jar tests, the results of which may not be directly applicable. However, they are at least indicative of performance. As an illustrative example, the procedures and results from the Manassas studies (8, 16) are utilized in the subsequent paragraphs.

Figure 7-2 shows the total soluble phosphorus as a function of the metal ion dose for the Manassas raw wastewater, for mixed liquor from a high rate activated sludge system, and for biologically stabilized final effluent. Since both the alum and ferric chloride can cause an alkalinity depletion and potential pH drop, pH values are also recorded. Superior performance for alum was found at all mass dosages with an apparent limiting residual of 1.0 mg/l as P for the raw wastewater and final effluent, and less than 0.1 mg/l as P for the aerator mixed liquor. A yellow color in the product water was noted with the higher iron dosages.

Replotting the aluminum data from Figure 7-2 in Figure 7-3 shows the influence of

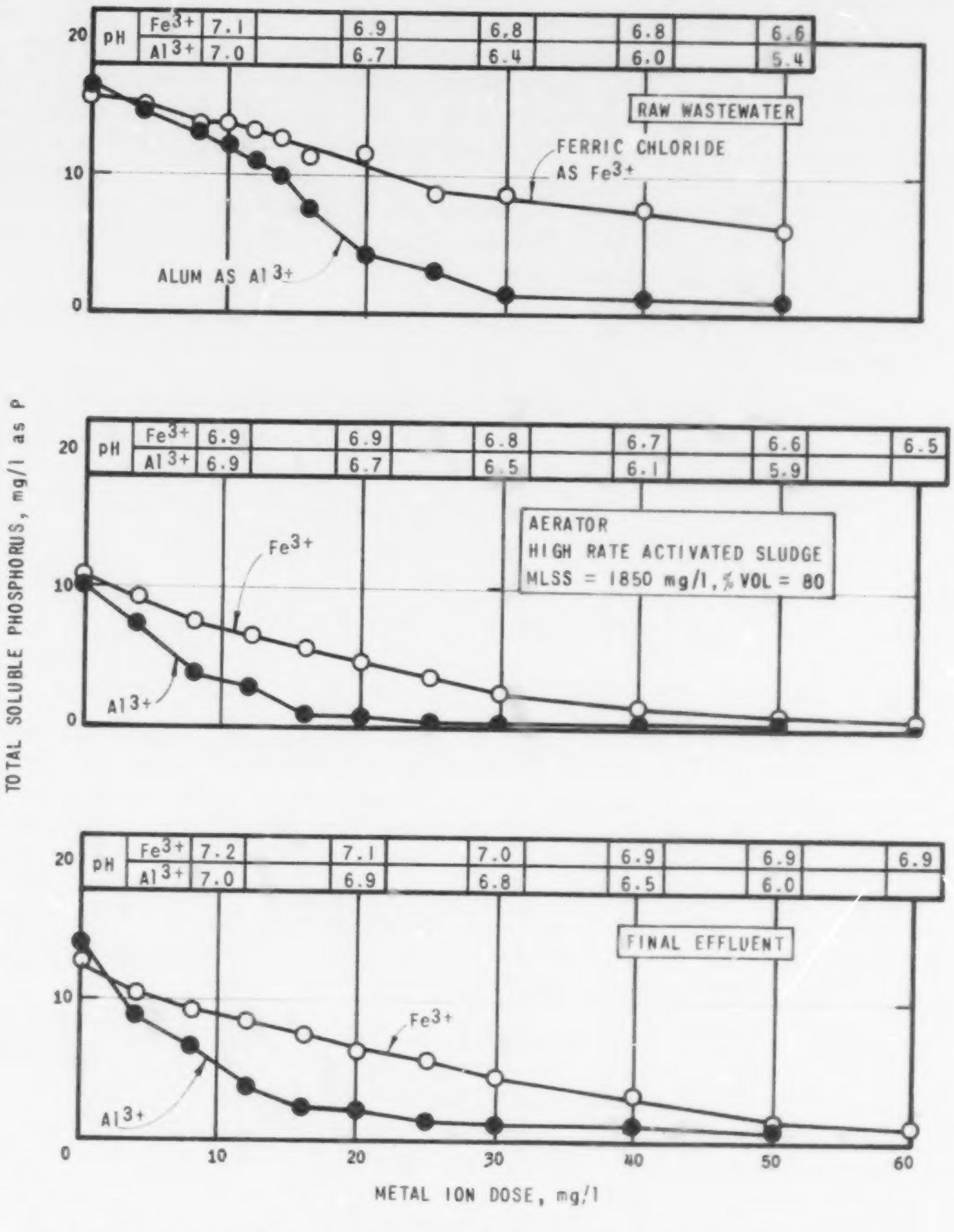

FIGURE 7-2 TOTAL SOLUBLE PHOSPHORUS RESIDUAL VERSUS METAL ION DOSE

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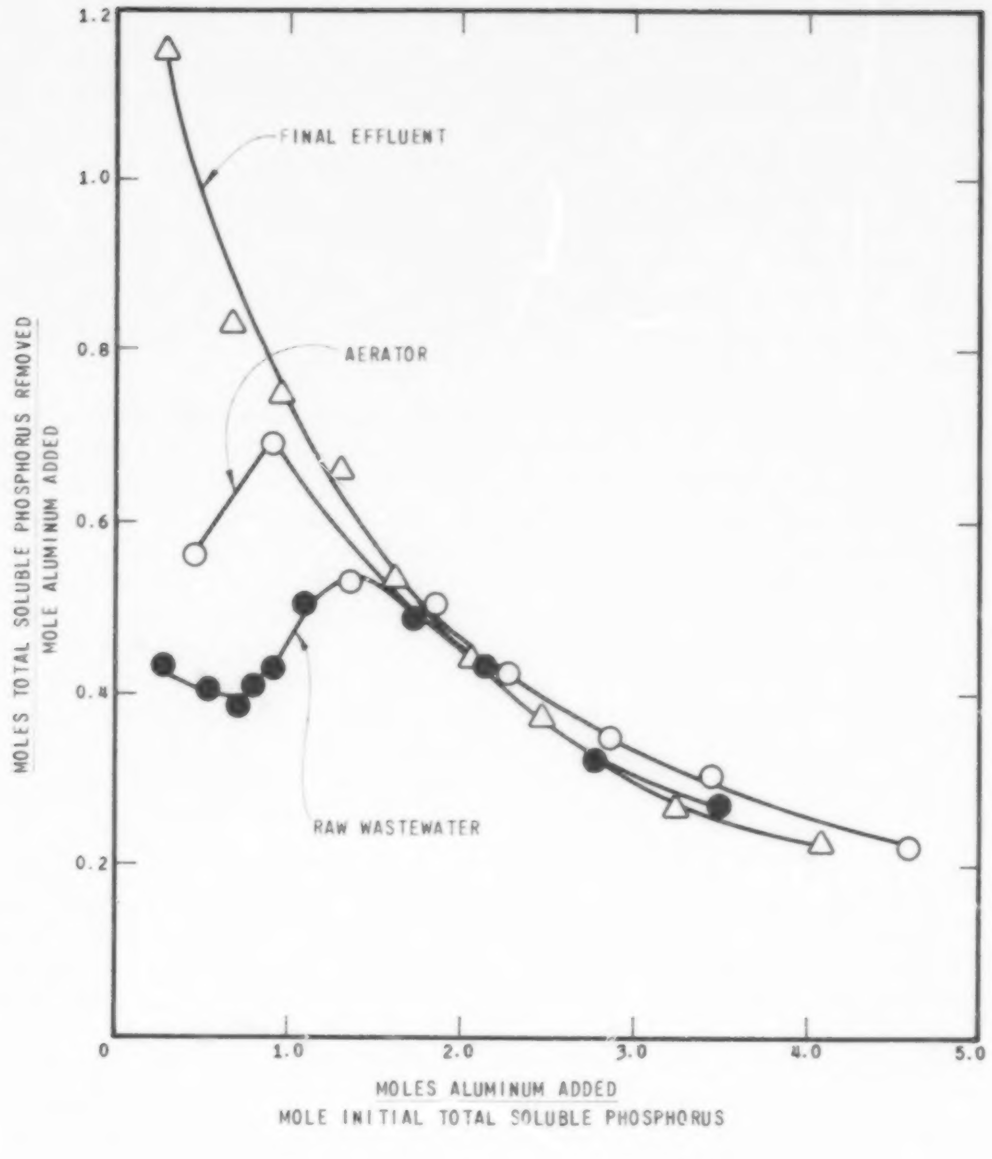

FIGURE 7-3 INFLUENCE OF POINT OF ADDITION ON PHOSPHORUS REMOVAL WITH ALUMINUM

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the point of addition on the molar effectiveness of the added aluminum. Similar but lower molar curves are found with the iron versus phosphorus residual data. The lag in phosphorus removal effectiveness encountered with the raw wastewater and aerator mixed liquor is thought to be due to the presence of unhydrolyzed phosphorus and the occurrence of competing side reactions for the metal ion. Aluminum and iron are known to form insoluble heavy salts with certain hydrophilic colloids such as detergents and soluble proteins and their degradation products (17), all of which would be present in raw and partially stabilized wastewater. These curves indicate that at molar ratios of less than 1:1, there is an obvious advantage in phosphorus removal derived by adding the chemical to a well stabilized liquid, whereas at molar ratios above 1.5:1, enough metal ion has been added to satisfy the competing side reactions and all points of addition are equally effective. The data indicate that very low phosphorus residuals are possible in a combined chemical-biological system and that the ideal point of addition would be just before final solids-liquid separation to obtain maximum biological stabilization of the soluble liquid phase.

Figure 7-4 shows the influence of wastewater and process pH upon phosphorus removal effectiveness with and without mineral additions. In the upper graph of Figure 7-4, above pH 8, phosphorus removal is occurring naturally through the formation of an insoluble calcium-phosphate complex via the natural hardness of the wastewater. The lower graph of Figure 7-4 shows that phosphorus removal at any molar dosage of metal ion is actually controlled by pH dependent reactions, optimally around 6 and 5 for aluminum and iron, respectively. Since the optimum range for aerobic oxidation of carbonaceous compounds is about pH 6 to 8, it can be seen that there is an opportunity for the simultaneous optimization of carbon oxidation and phosphorus removal by aluminum addition. However, if nitrification is a treatment goal there exists a conflict and, in turn, a compromise in pH goals. Diminished nitrification performance has been reported when alum additions have depleted the wastewater's alkalinity to a point where the pH was near optimum for phosphorus removal (9).

Table 7-1 provides a further comparison of representative sources of aluminum and iron. It should be noted that dissolved solids or extraneous ions are introduced and in some circumstances these ions may be undesirable (18). Alum and $FeCl_3$ cause an alkalinity loss, and without buffering capacity in the wastewater, a substantial pH reduction may occur. This can be compensated for by adding a source of alkalinity, such as lime, or by using aluminate. Treatment performance can deteriorate dramatically if the system pH is allowed to fall much below optimum (5). The designer should be aware that periods of heavy storm water infiltration will substantially dilute the normal alkalinity of the wastewater, and if operation is carried out at a pH of 6, only 20 to 40 mg/l of residual $CaCO_3$ buffering alkalinity may remain (8). The bottom part of Table 7-1 shows a comparison of the metal salts on a mole per mole basis, assuming a mole of metal ion precipitates one mole of phosphorus. In practice this never occurs. On this basis the possible cost advantage and alkalinity advantage with iron is lost and the aluminum salts show a potential for less additional solids production.

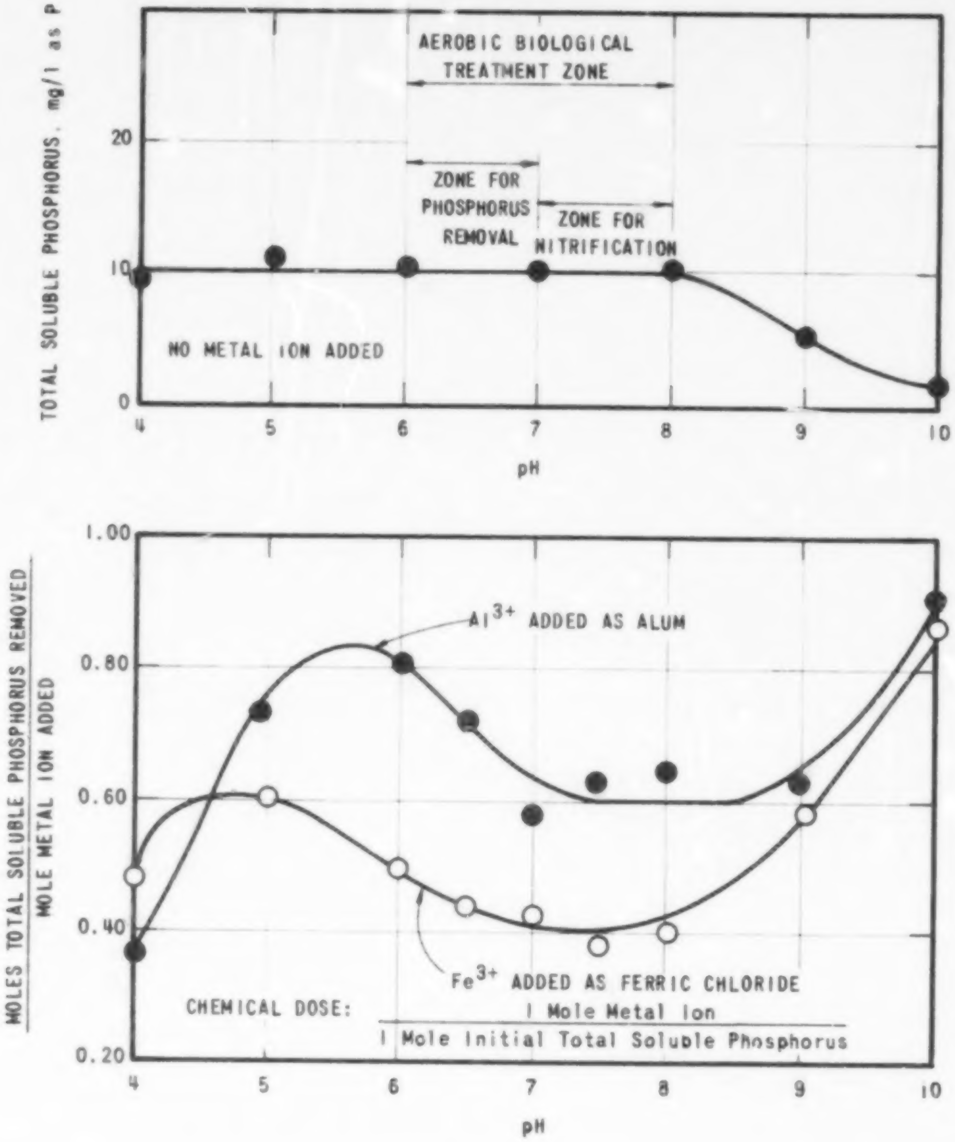

FIGURE 7-4 EFFECT OF pH ON PHOSPHORUS REMOVAL FROM A FINAL EFFLUENT

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Table 7-1

METAL SALT COMPARISON

		Metal Salt	
	Liquid Sodium Aluminate (Al^{3+})	Liquid Alum (Al^{3+})	Ferric Chloride (Fe^{3+})
	(per lb of Metal Ion)		
Bulk Delivered Price at Manassas, $	0.69	0.32	0.23
Extraneous Ion Introduced, lb	Na^+ 0.85	SO_4^{2-} 5.4	Cl^- 1.9
Maximum Alkalinity Loss as $CaCO_3$, lb	None	5.5	2.7
	(Per lb of Phosphorus at a Mole Dosage of 1:1)		
Bulk Delivered Price at Manassas, $	0.60	0.28	0.41
Extraneous Ion Introduced, lb	Na^+ 0.74	SO_4^{2-} 4.7	Cl^- 3.4
Maximum Alkalinity Loss as $CaCO_3$, lb	None	4.8	4.8
Phosphorus Removal of Metal Ion, lb	0.87	0.87	1.8
Additional Solids Production, lb	$AlPO_4$ 3.9	$AlPO_4$ 3.9	$FePO_4$ 4.9

Several other points should be considered in mineral selection. First, in the process effluent the soluble Al, less than 0.5 mg/l (16), is apparently much lower than found with iron additions, 6 mg/l Fe reported at Cleveland (19). Second, alum adds a divalent ion (SO_4^{2-}) which has a proven coagulation benefit (20). Third, both Mulbarger (8) and Long, et al. (9), report improved product water clarity when using alum in comparison to aluminate.

Chemicals can be added with confidence in the activated sludge process just before solids-liquid separation, assuming adequate dispersal. Studies by Allied Chemical Company (21) in a 25,000 gpd pilot plant indicated that the addition of alum at the seven-eights point in the aeration basin, along with anionic polymer in the transfer line, provided the optimum phosphorus removal, coagulation and settling. Polymer dosages were found to be optimum in the 0.2 to 0.4 mg/l range. An average Al:P weight ratio of 2.1:1 was required to reduce the average influent total phosphorus of 3.6 mg/l to 0.45 mg/l in the effluent.

It should be noted that utilization of activated sludge aeration tanks for chemical mixing and flocculation is a compromise. Typically, a diffused aeration system will have a velocity gradient of over 100 ft/second/ft and the velocity gradient of a mechanical aeration system will be much higher. Water treatment experience has shown that velocity gradients greater than 75 ft/second/ft will result in the onset of chemical floc disintegration (22). Undoubtedly, some chemical floc as well as biological floc disintegration does occur in activated sludge tanks and, because of solids recycle, is unavoidable. Floc disintegration and mineral overdose both lead to product water quality degradation. In new design situations it may be advantageous to provide a brief period of gentle solids agitation prior to solids separation to promote reflocculation of the chemical-biological solids. This can be done by gentle air agitation, a flocculation tank, or a flocculator-clarifier. The designer should also provide the capability of adding organic polymer, if needed. Nalco 676 and Atlas 2A2, slightly anionic organic polyelectrolytes, have been successfully used at dosages between 0.3 and 0.4 mg/l (8).

7.3 Performance and Optimization

7.3.1 Phosphorus Removal

At molar dosages of 1 to 2, metal ion to phosphorus, soluble phosphorus residuals of 0.3 to 3.0 mg/l as P are frequently encountered in the literature. Interpretation of some of the data is complicated by failures to establish biological equilibrium [90% equilibrium values are reached at approximately 2.2 times the operating cell retention time (16)] prior to changing variables or initiating an investigative run, and failure to consider and report pH.

Detailed data from iron additions to activated sludge systems are limited in the literature and consequently the more extensively available data on aluminum addition provide a better source of definitive design information. However, a few investigations of iron additions to secondary treatment processes have been conducted on the pilot and full plant scale.

Pilot plant studies by the Detroit Metro Water Department, in which several types of activated sludge systems were investigated, indicated that the most effective total phosphate reduction was provided by the conventional activated sludge process with iron ($FeCl_2$ derived from pickle liquor) fed to the primary effluent (23). The majority of the test runs exhibited removals varying from 75 to 85% for inorganic phosphate concentrations ranging from 1.9 to 26.6 mg/l as P in the raw wastewater. Phosphate removals by the activated sludge process alone (no iron fed) fell principally in the 55 to 70% range. It was found that, in general, the addition of iron to the conventional activated sludge process provided a treatment system capable of consistently high phosphorus reduction. The step-feed, activated sludge process, with the addition of iron or iron, polymer and caustic soda, also provided consistent total phosphorus reduction. It was concluded that the full-scale plant should be designed for the activated sludge process, arranged for both conventional and step-feed configurations, and that provision be made for phosphorus removal, utilizing injection of steel pickle liquor ($FeCl_2$).

A pilot plant (5,760 gpd) at Warren, Michigan (24) was designed to simulate the existing 36 mgd activated sludge plant, with the addition of one sand filter and one multimedia filter operated in parallel after the final clarifiers. Ferric chloride was fed to the effluent of the aeration basins to determine whether or not this method would meet the State-required 80% phosphorus removal. Ferric chloride addition at a ratio of 1.5 mg/l of Fe to 1.0 mg/l of soluble PO_4^{3-} reduced the average total phosphorus concentration from 7.0 mg/l to 1.2 mg/l in the effluent. This amounted to a reduction of 83%.

Leary, et al., of the Milwaukee Sewerage Commission (25), conducted a one-year plant-scale study of the use of waste pickle liquor ($FeSO_4$) to enhance phosphorus removal. The Milwaukee Jones Island Plant consists of a single primary facility followed by two parallel activated sludge plants. Pickle liquor was employed for phosphorus removal at the 115 mgd East plant during the test period. The 85 mgd West plant served as a control. Pickle liquor feed was equivalent to an average Fe^{2+} dosage of 9.4 mg/l. The pickle liquor was added before the aeration tank about 55 feet downstream from the point of addition of return sludge. Based on a 1970 average total phosphorus concentration of 8.2 mg/l P in the screened sewage, the East Side plant, with iron addition, accomplished 91.3% phosphorus removal while the control West plant removed 83.1%. Effluent phosphorus averaged 0.7 mg/l at the East plant and 1.4 mg/l at the control plant. The return sludge phosphorous concentration was 2.29% P in the control plant and 2.61% P in the East (test) plant. The iron content of the return sludge was increased from 1.86 to 5.08%.

The pickle liquor addition increased the East plant effluent SO_4^{2-} concentration from 123 to 145 mg/l and decreased the alkalinity from 213 to 169 mg/l as $CaCO_3$. Yearly average pH values were 7.0 and 7.1, respectively, for the East and West plants. The free acid in the pickle liquor used (two sources) ranged from 2.1 to 9.3% H_2SO_4.

Comparison of the BOD, COD, and suspended solids removal efficiencies of the West and East plants indicated that the addition of unneutralized pickle liquor did not adversely affect purification, and it did not cause problems with the plant physical facilities. The results of this one-year study indicated that the use of waste sulfuric acid pickle liquor as a source of iron for phosphorus precipitation and removal was practical and effective in maintaining low plant effluent phosphorus residuals.

In studies at the University of Missouri at Rolla (26), alum and aluminate at various times were fed to laboratory activated sludge units and compared to a control unit. The alum was more effective, as shown on Figure 7-5, than aluminate, based on the Al:P mole ratio. In the activated sludge units which were dosed with aluminum, the volatile mixed liquor suspended solids dropped to 50 to 60% of the total mixed liquor suspended solids, compared to a value of 80% in the control. The aluminate units had higher pH, but all pH values were in the range of 7.7 to 8.2.

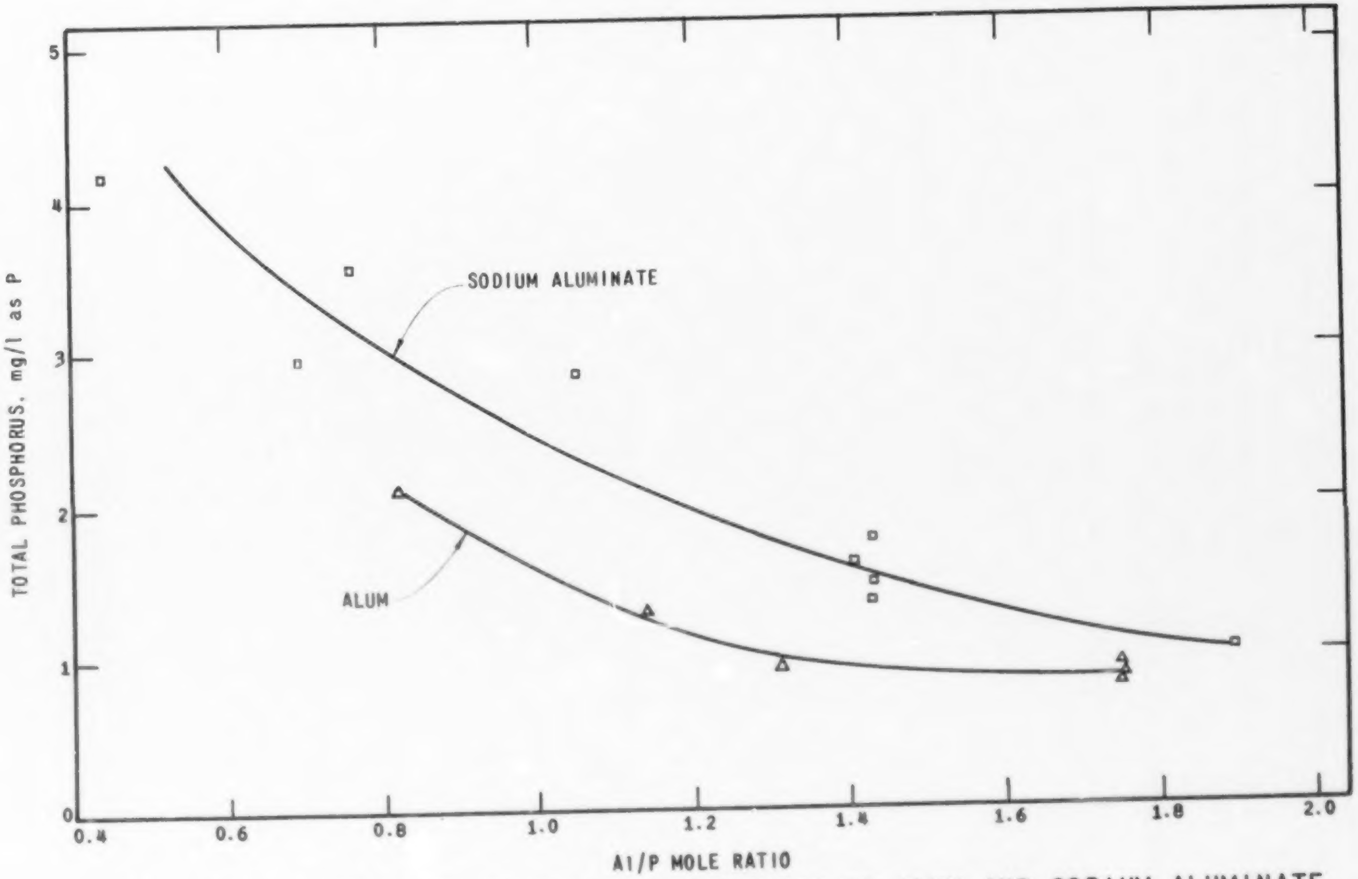

FIGURE 7-5 COMPARISON OF THE EFFECTIVENESS OF ALUM AND SODIUM ALUMINATE ON PHOSPHORUS REMOVAL

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Full-scale studies were conducted at Pennsylvania State University on the removal of phosphorus by addition of aluminum to a 2.0 mgd activated sludge plant. These studies began in 1969 and continued through 1970 (9). In the first phase of the work, alum and sodium aluminate were compared with respect to their ability to precipitate phosphorus. To prevent biological upset due to mineral solids displacing biological solids in the mixed liquor, the volatile suspended solids concentration was held at the same level. Since the volatile fraction of the total mixed liquor suspended solids decreased due to mineral solids buildup, this required the maintenance of a higher total suspended solids concentration in the mixed liquor.

The data shown by Table 7-2 indicate that alum was slightly superior to sodium aluminate in removing phosphorus from the moderately alkaline wastewater. The best point of addition for alum was found to be at the effluent channel of the aeration basin which carries mixed liquor to the final settling basin. Addition of alum both at the inlet end of the aerator and at a point about two-thirds of the way along the basin from the inlet resulted in lower removal efficiencies. Alum addition ahead of aeration produced a cloudy effluent.

Table 7-2

COMPARISON OF USE OF ALUM AND
SODIUM ALUMINATE AT PENNSYLVANIA STATE UNIVERSITY

Parameter	Chemical*			
	Alum		Sodium Aluminate	
	Influent	Effluent	Influent	Effluent
Suspended Solids, mg/l	134	19	94	25
BOD_5, mg/l	87	6	50	7
COD, mg/l	244	31	160	36
Phosphorus as P, mg/l				
Filtered Ortho	6.4	0.28	5.4	0.51
Filtered Total	7.6	0.27	6.9	0.57
Unfiltered Ortho	10.4	1.17	6.2	1.14
Unfiltered Total	12.4	1.48	8.6	1.53
Alkalinity as $CaCO_3$, mg/l			156	103
pH	7.4	7.1	7.15	7.25
Alum Dosage as $Al_2(SO_4)_3 \cdot 14H_2O$, mg/l	160			
$Na_2Al_2O_4$ dosage, mg/l			44	
Al:P Ratio **, weight	2.04:1		2.04:1	
SO_4^{2-}, mg/l	24	131		

*Chemicals added at the effluent end of the aeration tank.

**Al:P ratio based on expression of phosphorus in the filtered ortho form.

Aluminate addition was found to be most effective at a point in the aeration basin near the outlet. Aluminate addition either before or after aeration produced a more turbid effluent.

Sludge production in the studies comparing alum and aluminate showed that the volume of sludge produced per million gallons of wastewater treated was about the same for both coagulants. However, the weight of solids produced was considerably less for aluminate than alum.

Because alum was found to be superior to aluminate in removing phosphorus, further study of alum was performed for a period of one year. One-half of the activated sludge plant was used in the alum study and the other half was operated without chemical addition for comparison. Table 7-3 shows average results for operation at rates equal to or less than the plant design capacity.

Table 7-3

COMPARISON OF ALUM ADDITION TO THE AERATOR AND CONVENTIONAL ACTIVATED SLUDGE AT PENNSYLVANIA STATE UNIVERSITY FOR FLOWS NOT EXCEEDING THE DESIGN CAPACITY OF THE PLANT

Parameter	Influent	Effluent with Alum Addition*	Effluent without Alum Addition
Suspended Solids, mg/l	110	22	26
BOD_5, mg/l	71	9	13
COD, mg/l	172	55	68
Phosphorus as P, mg/l			
Filtered Ortho	6.7	0.28	6.7
Filtered Total	6.3	0.36	6.7
Unfiltered Ortho	8.9	1.15	7.2
Unfiltered Total	10.0	1.41	7.3
Alkalinity as $CaCO_3$, mg/l	168	80	120
pH	7.6	6.75	7.35
Al:P Ratio**, Weight		2.97:1	

*Alum added to effluent of aeration tank. Values shown in this column are means for the period of study.

**The Al:P ratio is based on expression of phosphorus in the filtered ortho form.

Problems were experienced with loss of solids from the final basin during peak flow periods where alum treatment was being practiced. It was believed, however, that proper allowance for peak flows in design of such facilities would eliminate excessive solids loss. The effluent insoluble phosphorus concentration was found to correlate well with the effluent suspended solids concentration.

A 2,000 gpd pilot operation at the main wastewater treatment plant of the District of

Columbia was operated at a constant alum dose rate (27). The pilot plant is a four-stage step aeration process followed by secondary clarification. Alum was added to the wastewater of moderate alkalinity (100 to 150 mg/l as $CaCO_3$) at the fourth stage of aeration using an Al:P weight ratio of 2:1 (molar ratio 2.3:1), based on an influent phosphorus concentration of 8.15 mg/l as P. Mixed liquor suspended solids averaged 6,300 mg/l with a volatile content of 60% and a cell retention time of about 5 days. Total phosphorus residuals averaged 1.9 mg/l as P (78% removal) in the clarifier effluent.

Figure 7-6 shows the chemical phosphorus removal found for the Manassas high rate activated sludge plant with aluminate (16) and alum (8) additions at slightly alkaline pH values. Phosphorus concentrations for Figure 7-6 were corrected for biological removal. The Manassas Pilot Plant, shown in Figure 7-7, is a 0.2 mgd three-stage (nitrification-denitrification) activated sludge system operated at the Greater Manassas Sanitary District of Prince William County, Virginia. Table 7-4 shows the system performance for 200 determinations with alum addition. The reader is referred to the original publication for more detailed information.

Table 7-4

COMBINED CHEMICAL–HIGH RATE
ACTIVATED SLUDGE SYSTEM PERFORMANCE

Item	Min.	Percent Less Than					Max.
		10	20	50	80	90	
Primary Effluent							
COD, mg/l	27	64	86	152	227	250	324
P, mg/l	2.2	3.8	5.0	8.4	11.4	12.8	18.5
BOD_5:COD*	0.26			0.43			0.56
High Rate Activated Sludge Effluent							
Soluble Phase							
COD, mg/l	13	23	26	33	41	48	79
P, mg/l	0.2	0.8	1.0	2.0	3.0	3.6	6.8
BOD_5:COD*	0.14			0.28			0.36
Solid Phase							
Suspended Solids, mg/l	0	10	14	21	36	44	196
BOD_5:SS*	0.10			0.23			0.31
COD:SS*	0.43			0.58			0.80
P:SS*	0.039			0.051			0.063
Suspended Solids After Polymer Addition,** mg/l	0	0	1	2	8	14	18
Suspended Solids After Mixed Media Filtration**, mg/l	0	0	0	0	2	5	12

*Ratio of minimum, median, and maximum monthly averages for each parameter
**94 determinations

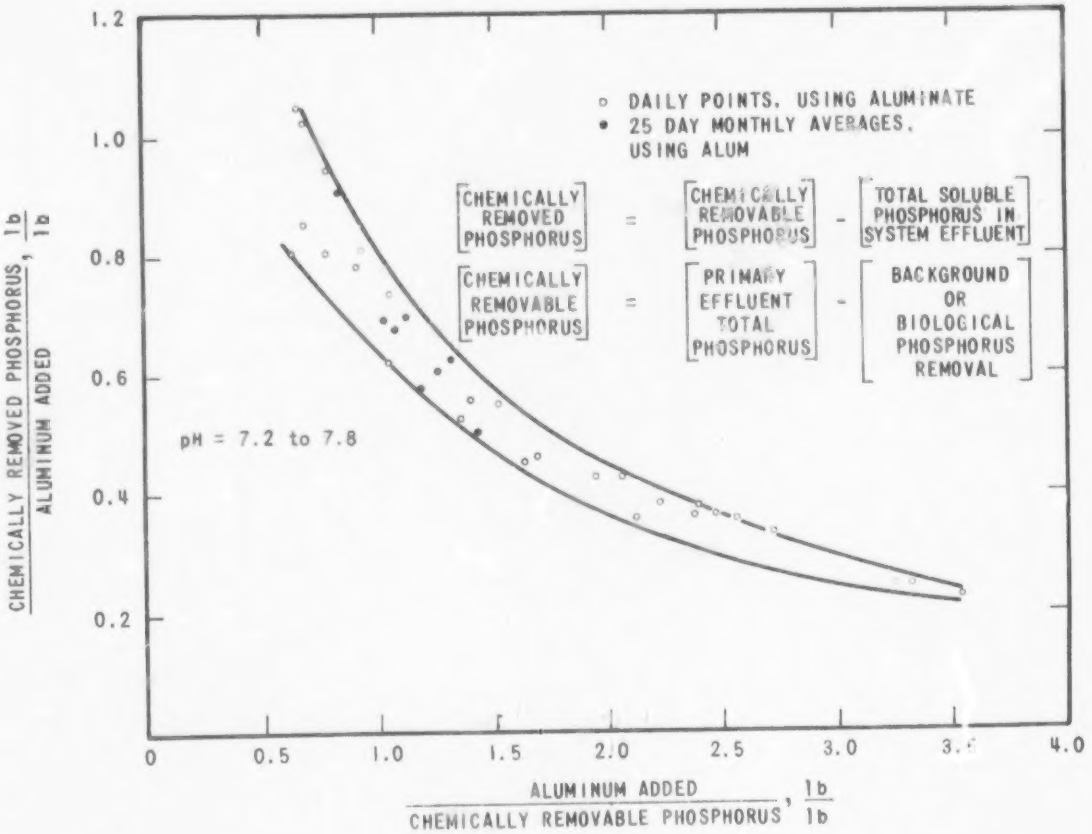

FIGURE 7-6 CHEMICAL PHOSPHORUS REMOVAL ENVELOPE:
HIGH RATE ACTIVATED SLUDGE SYSTEM

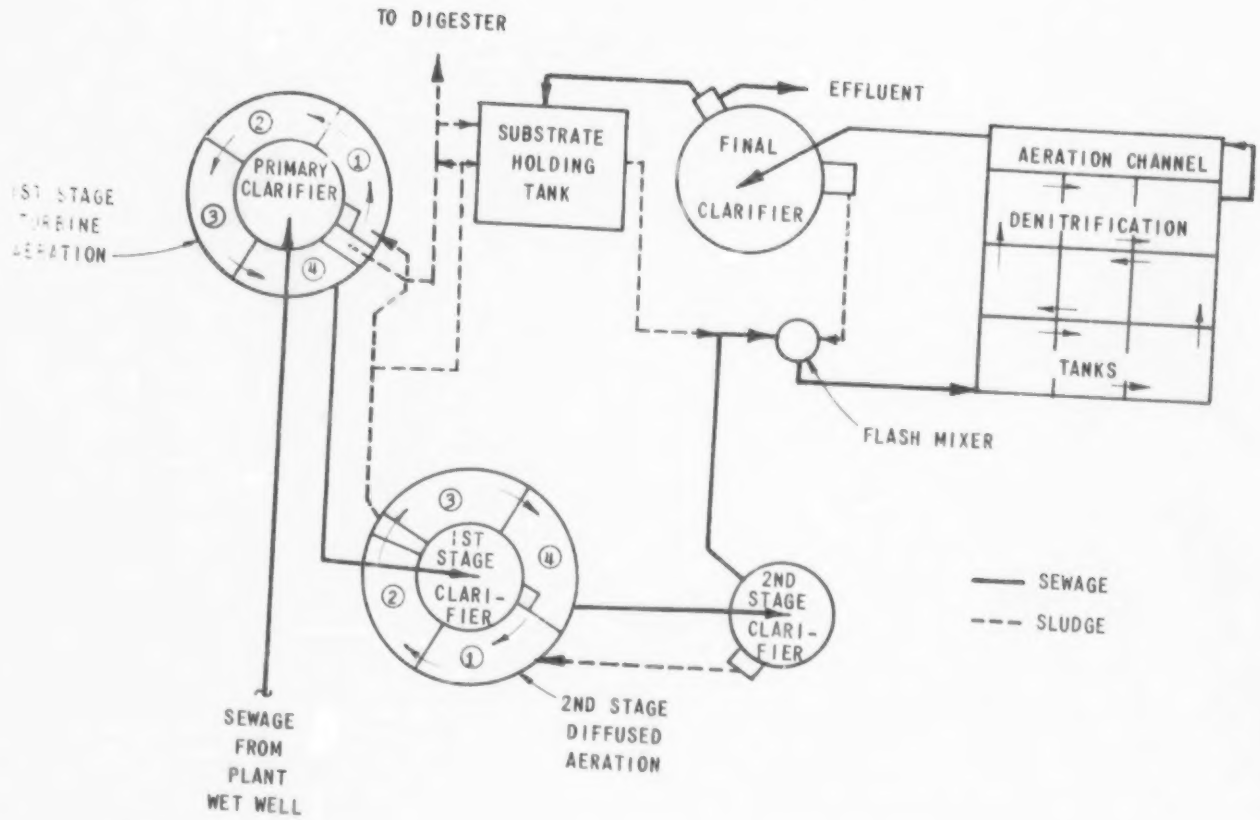
TO DIGESTER
EFFLUENT
PRIMARY CLARIFIER
1ST STAGE TURBINE AERATION
SUBSTRATE HOLDING TANK
FINAL CLARIFIER
AERATION CHANNEL
DENITRIFICATION
TANKS
FLASH MIXER
1ST STAGE CLARIFIER
2ND STAGE CLARIFIER
2ND STAGE DIFFUSED AERATION
SEWAGE FROM PLANT WET WELL
① ② ③ ④
— SEWAGE
--- SLUDGE

FIGURE 7-7 0.2 mgd BIOLOGICAL NITRIFICATION - DENITRIFICATION PILOT PLANT
Greater Manassas Sanitary District
Prince William County, Virginia

Optimization of phosphorus removals with mineral additions can be done two ways: by pH adjustment to the point of minimum solubility (pH 6 to 6.5, residual alkalinity of 20 to 40 mg/l as $CaCO_3$) and by splitting the chemical dose to separate parts of the treatment system (8). Optimization maximizes the efficient use of chemical, minimizes additional solids production, and produces soluble phosphorus residuals of less than 0.5 mg/l. Hais, et al. (27), indicated, using step aeration with alum addition, that above pH 6.6 a rapid increase in the effluent phosphorus residual occurred with increasing pH. However, operation with a mixed liquor pH of less than 6.2 caused an upset in the activated sludge process. It was concluded that optimum performance (0.6 mg/l effluent phosphorus residual) was achieved by maintaining a final stage pH between 6.3 and 6.6 at an Al:P weight ratio of 2:1. Mulbarger (8) showed the capability of pH adjustment with a denitrification system and demonstrated the advantages of split chemical treatment in a multi-sludge system. Results from the pH adjusted system are shown in Figure 7-8. Long, et al. (9), found a natural pH adjustment with high doses of alum in a poorly buffered wastewater and report phosphorus residuals below 0.5 mg/l at Al:P weight ratios of from 2:1 to 3:1.

Organism activity and type (9) as well as the completeness of a chemical reaction affect phosphorus removal and are influenced by the temperature. No temperature effect has been reported, however, for the rate of chemical reaction (13) operating between 10 and 20° C.

Mulbarger (8, 16) notes that the presence or absence of nonsettleable effluent suspended solids from a combined chemical-biological system is more dependent upon the ratio of net volatile solids produced to aluminum added than upon pH, exceeding an aluminum to phosphorus ratio, or exceeding a given aluminum dosage. It is recommended that this ratio not drop below 3 to 5. It is believed that the biological volatile solids provide exchange and/or sorption sites of limited capacity for the precipitated aluminum-hydroxy-phosphate complex, and that these natural polymers aid in clarification. Thus, the more biological solids produced from the system the greater the aluminum dosage that can be used without effluent suspended solids problems.

When designing the phosphorus removal system, the designer should be cognizant of the phosphorus levels in the effluent suspended solids. There was a nominal 1.0 mg/l of phosphorus in the effluent suspended solids from the high rate system at Manassas (8). After pH adjustment and addition of 0.3 to 0.4 mg/l of anionic polymer (Nalco 676), effluent suspended solids were reduced to 2 mg/l. Consequently, the effluent total phosphorus was also reduced. Polymer performance was also reported to be influenced by the pH (8). If total phosphorus residuals of less than 0.5 mg/l as P are required, a multimedia filtration system is recommended. Dual and tri-media filtration both provide effective tertiary solids separation. Hais, et al. (27) showed that tri-media filtration consistently removed between 5 and 10% more suspended material than did the dual-media filter. Filter runs between 24 and 32 hours were regularly maintained. Tri-media filtration performance was reported to be superior to dual-media filtration at Manassas (8).

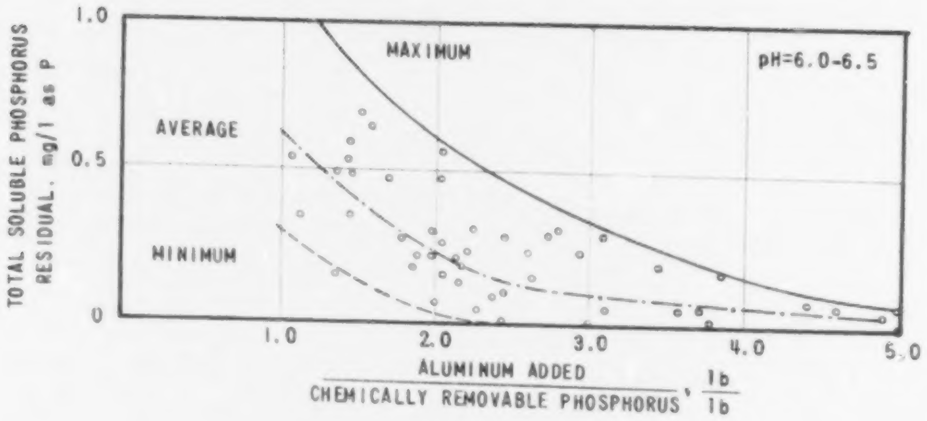

FIGURE 7-8 EFFECT OF ALUMINUM ADDITION WITH pH ADJUSTMENT ON TOTAL SOLUBLE PHOSPHORUS RESIDUAL

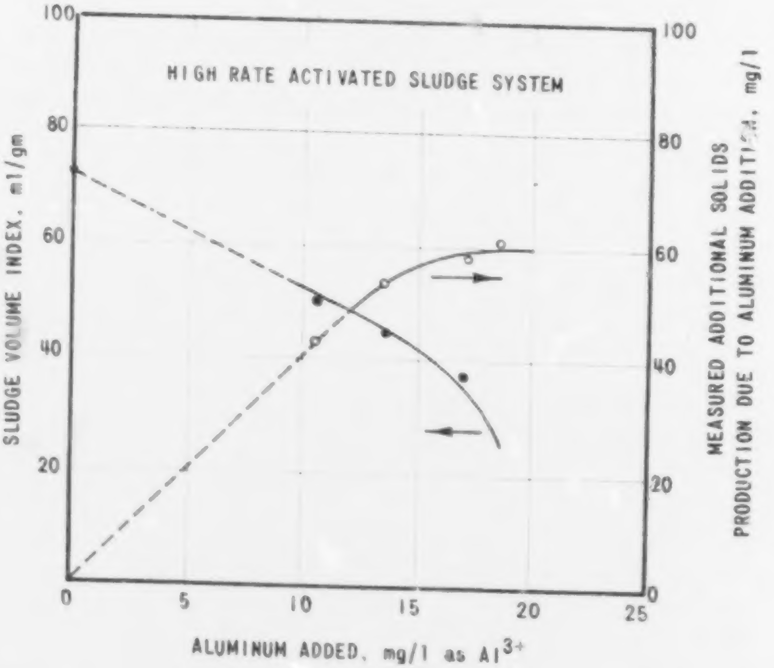

FIGURE 7-9 CHANGE IN SLUDGE VOLUME INDEX AND ADDITIONAL SOLIDS PRODUCTION DUE TO ALUMINUM ADDITION

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7.3.2 Sludge Characteristics

Long, et al. (9), indicated in studies conducted at Pennsylvania State University that sludge production in the chemical-biological process was significantly greater than in biological treatment only. This is to be expected because of the sludge produced by mineral addition. For a 1 year alum test period, the amount of waste sludge from the chemical-biological process was 2.1 times by weight that from the biological process alone. On the basis of volatile solids, the ratio was 1.56:1. It should be pointed out, however, that high alum doses (mole ratios of Al:P of 2:1 or above) were used in the study and that the sludge age of the undosed control was greater than that for the dosed system.

Figure 7-9 shows the changes found at Manassas (16) in sludge density and additional solids production with aluminate additions. During these studies the average weekly primary effluent phosphorus concentration ranged between 9.3 and 13.1 mg/l. From this figure it can be concluded that up to stoichiometric requirements, additional solids production will be about 4 lb/lb Al^{3+}. Above stoichiometric requirements, production of additional solids will drop off. This observation can be explained by the predominant formation of $AlPO_4$ (molecular weight, 122) at less than stoichiometric Al additions and formation of $Al(OH)_3$ (molecular weight, 78) above stoichiometric Al additions. Figure 7-9 also shows an improvement in sludge density (which is inversely related to the sludge volume index) with increasing aluminum dosages. Whether or not this density improvement is enough to compensate for the additional solids that must be handled, and thereby eliminate the need for additional sludge disposal facilities, is dependent upon the original sludge density and the amount of phosphorus removed and aluminum added. Figure 7-10 shows actual plant data which indicate mineral additions do produce a dense, stable sludge.

Figure 7-10 also shows the change of the sludge settling characteristics as a function of the operating solids level (8). The reported sludge settling characteristics were determined in 1-liter graduates. Manassas experience also indicates that the settled sludge does have a tendency to classify (separation of biological and chemical fractions) although not to such an extent as to harm system performance. Therefore, sludge collection systems which have a tendency to classify sludges into heavy and light fractions, i.e., those that work with a hydrostatic or vacuum principle, probably should be avoided.

Sludges resulting from aluminum addition are amenable to anaerobic digestion (28). In conjunction with laboratory scale aluminum addition to activated sludge units at the University of Missouri at Rolla (26), laboratory scale anaerobic digesters were studied to determine the fate of the aluminum and phosphorus. The digesters were fed 70% activated sludge and 30% primary sludge. Detention time in the digesters was 15 days. The fate of the phosphorus in the digesters is shown in Figure 7-11. The phosphorus which had been precipitated with aluminum was concentrated in the digester sludge and not released to the supernatant. Tests confirmed that the aluminum was also retained in the digester sludge. Comparing the digesters which were fed aluminum

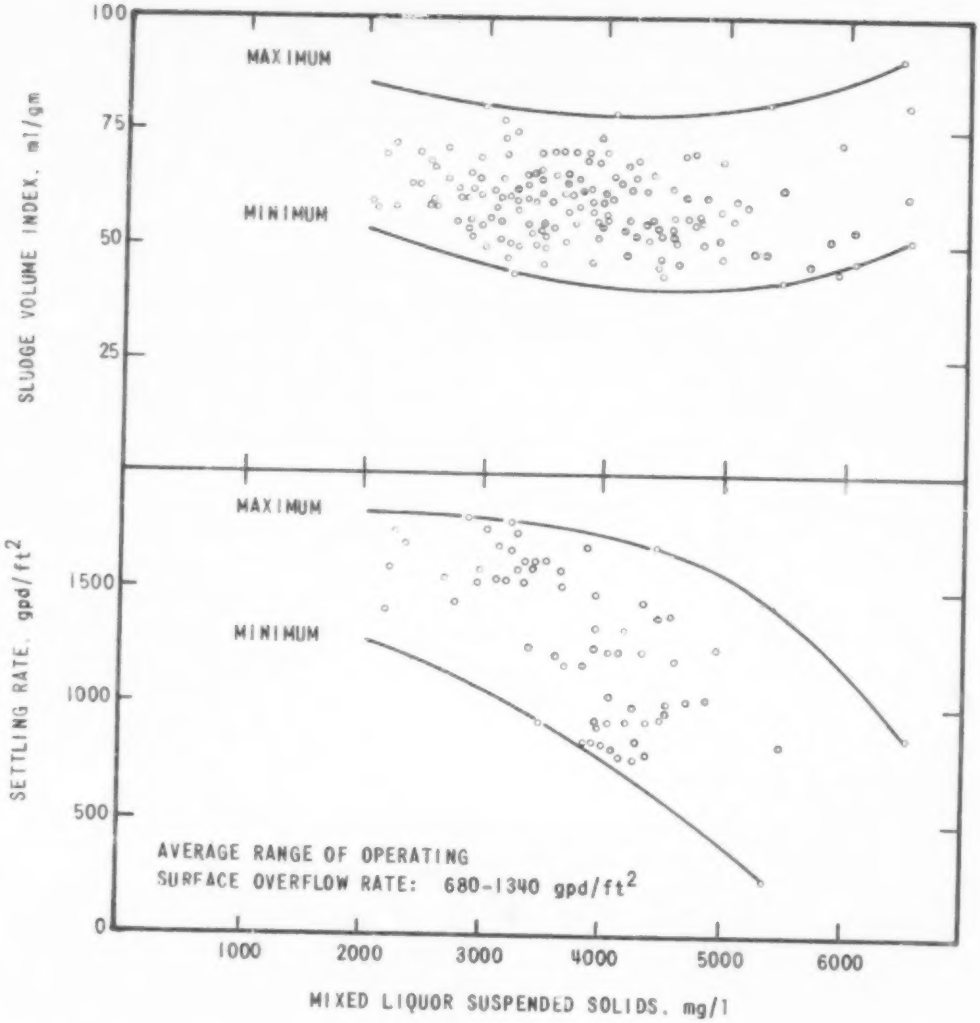

FIGURE 7-10 PHYSICAL CHARACTERISTICS OF HIGH RATE ACTIVATED SLUDGE WITH MINERAL ADDITION

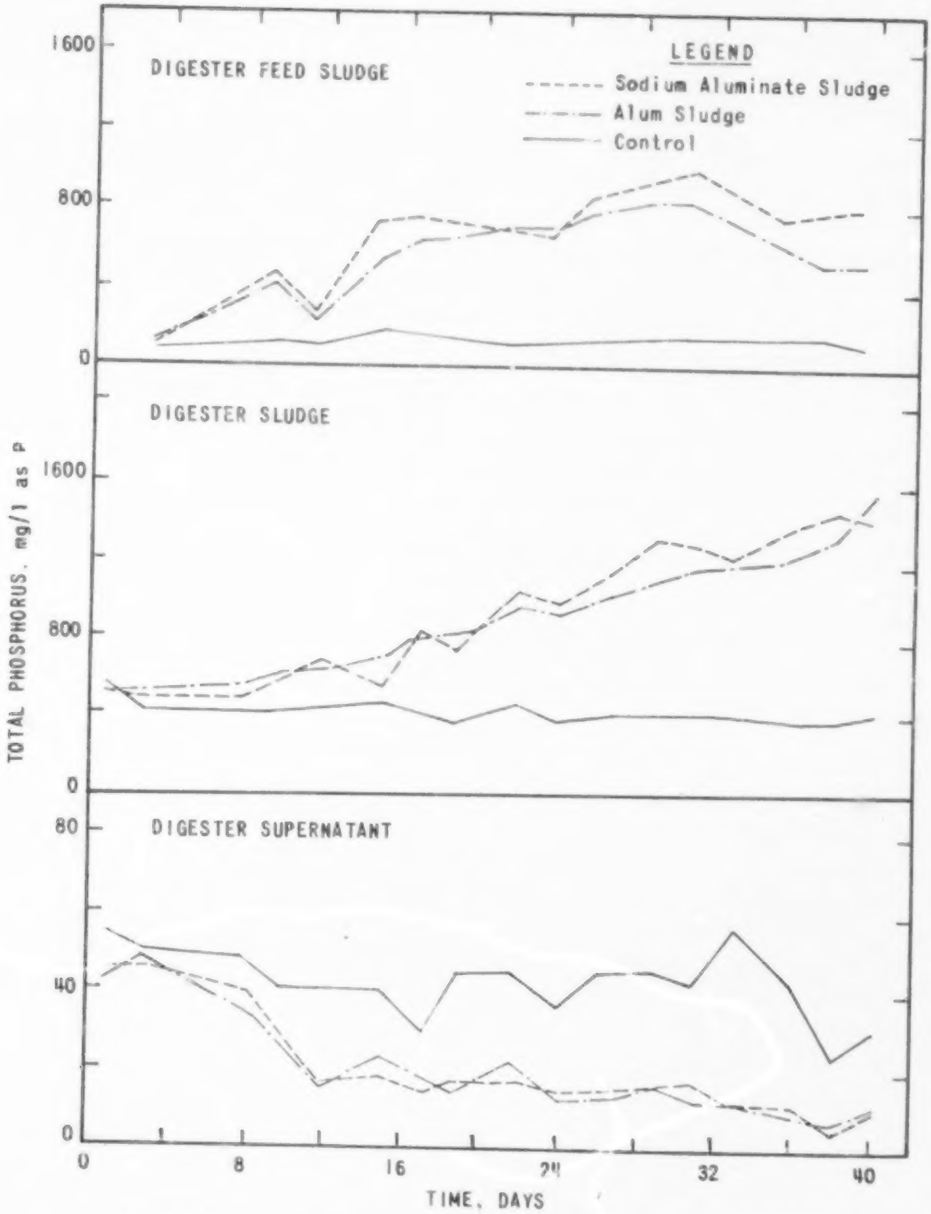

FIGURE 7-11 FATE OF ALUMINUM - PRECIPITATED PHOSPHORUS IN ANAEROBIC DIGESTION

(Reprinted with permission from Journal Water Pollution Control Federation in press, Washington D.C. 20016) sludge with a control digester, using the parameters of volatile acids and gas production, the digester performance was equivalent.

Experience at Manassas has found no resolubilization of phosphorus with the waste organic-chemical sludge in anaerobic digestion (16). Connell reports similar observations with iron additions (10).

At Manassas, comparative bench studies of settled high rate activated sludge dewaterability with and without aluminum additions showed that dry cake filter yields will substantially increase with aluminum additions, but there is a tendency to produce a wetter cake. Similar high yield rate observations after aluminum additions have been observed by Smith, et al. (29). This increase in dewaterability can probably be explained by the fact that with efficient chemical utilization, aluminum phosphate, not aluminum hydroxide, is the inorganic precipitate being dewatered.

Mulbarger (8) indicates that it would be reasonably conservative to design solids handling equipment for the organic solids loading on the basis of conventionally used criteria, assuming that the additional chemical solids that must be handled will be compensated for by the increase in sludge dewaterability.

No published data are available on the dewatering characteristics of iron-organic sludges.

Aluminum additions will result in the least amount of additional solids for ultimate disposal, probably 30 to 50% of that found with iron and 5 to 20% of that found with two-stage lime treatment.

Aluminum-organic sludges may be incinerated or spread on land. When the sludge is applied to the soil, leachates should not be a problem because of the natural phosphorus exchange capacity of soils and the low solubility of the aluminum phosphate precipitate. Preliminary tests by Dotson (30) showed no plant-inhibitory effect with applications of organic-aluminum sludges of up to 1 in. per week although some initial seed germination inhibition may be encountered at higher applications. Generally, it is felt that the soil building and fertilizing benefits derived from organic sludge application more than compensate for any deleterious effects from the inorganic aluminum sludge.

7.3.3 Costs

Power costs for combined chemical-biological treatment are negligible. Maintenance costs will increase somewhat due to the additional equipment. Capital expenditures for storage facilities, instrumentation, and equipment will be very low at larger facilities unless multi-media filtration is incorporated. Chemical costs will vary as a function of the incoming phosphorus concentration and the required phosphorus residual. Additional expenditures for pH control (alkalinity dependent) and solids control (polymer additions) may be necessary. See Chapter 10 for further information.

7.4 Process Design Examples

Four design example problems were selected for an arbitrary wastewater. Each illustrates specific points in combined chemical-biological wastewater treatment with alum additions for phosphorus removal. Results from the Manassas work are used for some of the calculations.

Assumptions:

a. Net biological solids production = 0.28 lb of volatile solids produced per lb of COD removed; the volatile solids contain 2% phosphorus; and the total solids average 80% volatile.

b. The biological system is to be designed with a minimum hydraulic contact time of 1 hour at the maximum 4 hour hydraulic peak. The ratio of maximum flow rate to average flow rate = 1.5. A cell retention time (CRT) of two days is selected.

c. Primary Effluent Characteristics:

COD = 200 mg/l
P = 10 mg/l
Alkalinity = 150 mg/l as $CaCO_3$

Find:

1. Effluent COD, BOD_5 and phosphorus
2. Total solids production
3. Alkalinity loss
4. Dissolved solids addition
5. Size of final clarifier, return and waste sludge pumping systems.
6. Chemical costs with the following delivered prices:

Alum	= \$0.32/lb Al^{3+}
Polymer	= \$1.50/lb polymer
Acid	= \$0.02/lb H_2SO_4
Lime	= \$0.01/lb $Ca(OH)_2$

Example No. 1: High rate system for most efficient chemical addition with pH above 7 and no optimization by pH adjustment.

1. **Effluent COD, BOD_5, and Phosphorus**
(See Table 7-4. Note: Soluble organics from a biological system vary as a function of the wastewater and dilution of the refractory fraction. Manassas values are used for illustration only.)

Effluent soluble COD $= \frac{200-152}{227-152}$ (41-33) + 33 [Interpolated from data of Table 7-4]

$= 5 + 33 = 38$ mg/l

Assume effluent suspended solids = 20 mg/l

Total effluent COD = 20 (0.58) + 38

= 12 + 38

= 50 mg/l

Effluent BOD_5 = 38 (0.28) + 20 (0.23)

= 10.6 + 4.6

= 15 mg/l

[Using BOD_5 ratios from Table 7-4]

Background phosphorus removal per lb COD removed

= (biological solids production ratio) (phosphorus content)

= (0.28) (0.02)

$= 0.0056 \frac{\text{lb phosphorus removed}}{\text{lb COD removed}}$

[The background phosphorus removal rate at Manassas was found to be approximately $0.01 \frac{\text{lb P Removed}}{\text{lb COD removed}}$ (8)]

Background phosphorus removal expressed as concentration

= (COD removal) (phosphorus removal per lb COD removed)

= (200-38) (0.0056)

= 0.9 mg/l

Chemically removable phosphorus = 10.0 - 0.9 = 9.1 mg/l

Use Al^{3+}/chemically removable phosphorus dose = 1.25 lb/lb

From Figure 7-6 find 0.6 lb of phosphorus removed per lb of aluminum added at the dosing ratio of 1.25

Aluminum dose = 1.25 (9.1) = 11.3 mg/l Al^{3+}

Chemical phosphorus removal = 0.6 (11.3) = 6.8 mg/l P

Soluble phosphorus residual = 9.1 - 6.8 = 2.3 mg/l

Total effluent phosphorus = 2.3 + 20 (0.05)

= 3.3 mg/l P

[P/SS ratio from Table 7-4]

The preceding calculations are useful from the standpoint that they show the pollutant fractions in both the soluble and solid phases. Often a remarkable treatment improvement can result from effluent solids control and minimization.

2. **Total Solids Production**

In calculating inorganic solids productions, dissolved solids additions, and alkalinity losses derived from alum additions, the designer should be aware of the two pH dependent competing reactions:

a. $Al_2(SO_4)_3 + 6HCO_3^- \longrightarrow 2Al(OH)_3 + 3\ SO_4^{2-} + 6CO_2$

(54 as Al^{3+}) (300 as $CaCO_3$) (156) (288)

or

5.5 lb of $CaCO_3$ alkalinity lost per lb of Al^{3+}

2.9 lb of $Al(OH)_3$ produced per lb of Al^{3+}

5.4 lb of SO_4^{2-} added per lb of Al^{3+}

b. $Al_2(SO_4)_3 + 2PO_4^{3-} \longrightarrow 2\ AlPO_4 + 3SO_4^{2-}$

(54 as Al^{3+}) (62 as P) (244)

or

4.5 lb of $AlPO_4$ produced per lb of Al^{3+}

3.9 lb of $AlPO_4$ produced per lb of P

0.87 lb of Al^{3+} used per lb of P

Because of phosphorus removal and aluminum uptake by "sorption" mechanisms, these preceding reactions do not completely describe observed phenomena but the estimates derived from them are close enough for engineering design.

Solids production due to $AlPO_4$ formation

$$= (6.8 \text{ mg/l phosphorus removed chemically}) \left(\frac{3.9 \text{ mg } AlPO_4}{\text{mg P}}\right)$$

$$= 27 \text{ mg/l } AlPO_4$$

Aluminum used in $AlPO_4$ formation

$$= (6.8 \text{ mg/l P}) \left(\frac{0.87 \text{ mg } Al^{3+}}{\text{mg P}}\right)$$

$$= 5.9 \text{ mg/l } Al^{3+}$$

Aluminum available for hydrolysis

$$= 11.3 - 5.9 = 5.4 \text{ mg/l } Al^{3+}$$

Solids production as $Al(OH)_3$

$$= 5.4 \text{ mg/l } Al^{3+} \left(2.9 \frac{\text{mg } Al(OH)_3}{\text{mg } Al^{3+}}\right) = 16 \text{ mg/l}$$

Total inorganic solids production per lb Al^{3+} added

$$= \frac{27 + 16}{11.3} = 3.8 \frac{\text{lb of solids}}{\text{lb of } Al^{3+} \text{ added}}$$

This result compares closely with the data in Figure 7-9. These calculations illustrate the point brought out in the text; i.e., the magnitude of solids production is a function of the proximity of aluminum dosage to the stoichiometric phosphorus requirement, assuming the pH is in the proper range.

Biological solids production:

$$= (\text{COD Removal}) (\text{Biological Solids Production Ratio}) \left(\frac{1}{\text{Volatile Fraction}}\right)$$

$$= (200\text{-}38) (0.28) \frac{1}{0.80}$$

$$= 57 \text{ mg/l SS}$$

Total Solids Production = Inorganic + Biological

= 27 + 16 + 57

= 100 mg/l SS

3. **Alkalinity Loss**

Maximum possible alkalinity loss due to aluminum hydrolysis

$$= 5.4 \ (5.5) = \underline{30 \text{ mg/l as } CaCO_3}$$

which gives $\frac{30}{11.3} = 2.7$ lb $CaCO_3$ loss/lb Al^{3+}

At an aluminum added to chemically removable phosphorus ratio of 1.16, an average alkalinity loss of 2.3 lb $CaCO_3$/lb Al^{3+} was measured at Manassas (8). Like solids production values, alkalinity losses are a function of the proximity of aluminum dose to stoichiometric phosphorus requirements and the reaction pH.

4. **Dissolved Solids Addition**

$$SO_4^{2-} = (11.3 \text{ mg } Al^{3+}/l) \left(5.4 \frac{\text{mg } SO_4^{2-}}{\text{mg } Al^{3+}}\right) = \underline{61 \text{ mg/l}}$$

Although dissolved solids are added to the system, the designer should remember that other soluble wastewater components (i.e., organics, phosphorus, alkalinity, etc.) are being removed. Prior to the addition of sulfuric acid for pH adjustment at Manassas, no increase in dissolved solids was measured on a primary to final effluent basis because of a compensating loss of soluble wastewater components (8).

5. **Final Clarifier and Pumping System**

Organic solids = 57 mg/l

Inorganic solids = 43 mg/l

Total = 100 mg/l

Solids lost in effluent = 20 mg/l

Solids accumulation = 80 mg/l

CRT = 2 days, and aeration tank detention at average flow

= 1.5 (1) = 1.5 hours

$$\text{MLSS in aerator} = \frac{(24)(2)}{1.5}(80) = 2{,}560 \text{ mg/l}$$

Since this is the final clarifier a conservative overflow rate should be utilized. From Figure 7-10 the minimum settling rate of 1,150 gpd/ft^2 was observed at a MLSS of 2,600 mg/l. To provide for peak flow, the clarifier should be designed for an overflow rate of 770 gpd/ft^2, or less. It should have a 12 ft effective water depth, and might be provided with a surface skimmer.

Maximum SVI at 2,600 mg/l MLSS = 82 ml/gm (Figure 7-10)

Approximate return sludge concentration at worst condition

$$= \frac{10^6}{\text{SVI}} = \frac{10^6}{82} = 12{,}200 \text{ mg/l}$$

Solids balance equation where R is sludge return rate

$$(Q_{max} + R)(\text{MLSS}) \leq (R)(\text{return sludge conc.})$$

or considering just the case of equality,

$$R = 1.5\, Q_{ave} \; \frac{2{,}600}{12{,}200 - 2{,}600}$$

$$R = 0.4\, Q_{ave}$$

(Actually, effective design would indicate a ±50% capability; i.e., 0.2 to 0.6 Q_{ave} capability)

Solids to be wasted per day = 80 mg/l at a minimum sludge solids concentration = 12,200 mg/l. Assume continuous wastage with separate pump.

$$W = Q_{ave} \frac{80}{12{,}200} = \underline{0.0065\ Q_{ave}}$$

(Again, effective design would indicate a ±50% capability; i.e., 0.003 to 0.01 Q_{ave} capability).

6. **Alum Cost**

$$= (11.3)(0.32)(8.34)(10^{-3}) = \underline{3.1¢/1{,}000\ \text{gal.}}$$

Additional Comments

No mention has been made of the effect of sidestream phosphorus contributions. With conventional solids handling and treatment processes, gravity and flotation thickening, filtration and centrifugation, aerobic and anaerobic digestion, and incineration, phosphorus recycle should not be a problem and any that occurs will largely be associated with the solids fraction of the minor process streams. However, processes that cause biological solids liquefaction and hydrolysis, such as heat treatment processes, will result in a substantial recycle of soluble phosphorus with a corresponding increase of metal ion requirements.

Example No. 2: Same system as Example No. 1 but with sulfuric acid addition for pH adjustment (residual alkalinity = 30 mg/l as $CaCO_3$) and polymer addition for solids control.

Detailed calculations are not shown for the problem since the techniques are essentially the same as in Example No. 1. Results are summarized, along with results from the other examples, in Table 7-5. Highlights from this problem are as follows:

1. COD, BOD_5, and suspended phosphorus concentrations in the effluent are substantially reduced by an assumed polymer dose of 0.3 mg/l. The chemically removed phosphorus was calculated for this case by multiplying the 0.6 lb P removed/lb Al^{3+} added, from Figure 7-6, by 1.33. The 1.33 was obtained from Figure 7-4, assuming an adjusted pH of about 6. The result was a phosphorus residual of 0.1 mg/l with an aluminum dose of 11.3 mg/l. But Figure 7-[illegible] showed from experimental data a nominal 0.5 mg/l residual phosphorus at this Al:P ratio and, therefore, the conservative residual of 0.5 mg/l soluble phosphorus was chosen.
2. Inorganic solids production consisted of 34 mg/l as $AlPO_4$ and 11 mg/l as $Al(OH)_3$ due to the preferential formation of $AlPO_4$ at an optimized pH.
3. The alkalinity loss due to hydrolysis was also reduced due to increased $AlPO_4$ formation. Using 96% strength sulfuric acid and assuming a residual alkalinity of 30 mg/l as $CaCO_3$, a calculated sulfuric acid requirement of 101 mg/l was found. The sequence of chemical addition is acid, alum, and polymer, with about 5 minutes of mixing at peak flows between each application.

Table 7-5

SUMMARY OF DESIGN CALCULATIONS

Item	Example 1	Example 2	Example 3	Example 4
pH Adjustment	No	Yes	Yes	Yes
Polymer Addition	No	Yes	Yes	Yes
Split Treatment*	No	No	No	Yes
Mixed Media Filtration	No	No	Yes	Yes
Total Aluminum Dose, mg/l	11.3	11.3	27.3	18.2
Effluent Quality, mg/l				
Suspended Solids	20	2	0	0
COD	50	39	38	–
BOD_5	15	11	11	–
Total Phosphorus	3.3	0.6	0.2	0.2
Total Soluble Phosphorus	2.3	0.5	0.2	0.2
Solids Production, mg/l				
Biological	57	57	57	57+
Chemical	43	45	92	66
Alkalinity Loss Due to Al^{3+} Addition mg/l as $CaCO_3$	30	21	109	59
SO_4^{2-} Increase Due to Al^{3+} Addition, mg/l	61	61	147	98
Alum Cost, ¢ / 1,000 gal.	3.1	3.1	7.4	4.9
Polymer Cost, ¢ / 1,000 gal.	0	0.3	0.3	0.3
Acid Cost, ¢ / 1,000 gal.	0	1.7	0	1.0
Total Cost, ¢ / 1,000 gal.	3.1	5.1	7.7	6.2
Clarifier Overflow Rate, gpd/ft^2	770	670	360	
Return Sludge Rate	0.4Q	0.6Q	0.9Q	
Waste Sludge Rate	0.0065Q	0.008Q	0.012Q	

*Split treatment here incorporates mineral addition as in Example No. 1 without pH adjustment, followed by mineral addition with pH adjustment at a later point.

4. Dissolved solids additions totaled 156 mg/l SO_4^{2-}; 61 mg/l due to the alum and 95 mg/l due to the acid.

5. Because of improved solids capture, the operating MLSS in the aerator increased to 3,200 mg/l necessitating a lower surface overflow rate in the clarifier, and greater return and waste sludge handling capability.

6. Chemical costs increased to 5.1¢/1,000 gal. with benefits in phosphorus, suspended solids, BOD_5 and COD removals.

Example 3: Assuming mixed media filtration provides complete solids control, what is the change in system calculations if a residual effluent phosphorus of 0.2 mg/l as P is to be provided?

Highlights from this problem are as follows:

1. Essentially no differences from Example No. 2 are found in effluent COD and BOD. However, the mixed media filter does provide positive effluent solids control in the event of system upset. An Al:P weight ratio of 3 (Figure 7-8) was used.

2. Inorganic solids production consisted of 35 mg/l as $AlPO_4$ and 57 mg/l as $Al(OH)_3$. These values are approximately double the total inorganic solids productions found in Examples No. 1 and 2.

3. The addition of alum caused an alkalinity consumption of 109 mg/l, which would deplete the background alkalinity of the wastewater to 41 mg/l as $CaCO_3$. If a substantially higher alum dosage had been required, all of the alkalinity might have been consumed, and addition of lime might have been necessary.

4. Dissolved solids addition consisted of 147 mg/l SO_4^{2-} from the alum.

5. Additional solids productions resulted in an operating solids level of 4,800 mg/l in the aerator, necessitating a lower surface overflow rate and higher return and waste sludge capabilities than in earlier examples.

6. Total chemical treatment costs are raised to 7.7¢/1,000 gal.

Example No. 4: Split treatment (chemical addition at more than one point) is utilized for a required effluent phosphorus concentration of 0.2 mg/l. Assume the first point of addition is as in Example No. 1. Further mineral addition will be made to the effluent in Example No. 1 either before filtration or during a second stage of biological treatment.

1. Changes in effluent COD and BOD are not calculated since this is dependent upon whether split treatment is practiced with just filtration, or whether another biological system is added. In a real situation it would be better to reduce the aluminum dose below that in Example No. 1 in the first stage system for more efficient chemical use. This was not done here for design simplicity. Another possibility is addition of part of the chemical dose in the primary settler.

   The Al^{3+} dose of 6.9 mg/l for the second addition was arrived at using Figure 7-8 for a pH adjusted system.

2. Additional inorganic solids production was 8 mg/l as $AlPO_4$ and 15 mg/l as $Al(OH)_3$.

3. The alkalinity depletion due to aluminum addition was calculated as 59 mg/l as $CaCO_3$. This would leave a residual alkalinity of 91 mg/l. Sixty mg/l as H_2SO_4 was added to reduce the alkalinity to 30 mg/l. Again, in a real situation acid might not have to be added to this wastewater.

4. Dissolved solids additions were calculated as 98 mg/l SO_4^{2-} due to alum addition and 59 mg/l SO_4^{2-} due to acid addition.

5. Clarifiers were not sized in this example. If two biological processes were used in series, the final clarifier of the first process could be designed with an overflow rate from the middle of the performance curve in Figure 7-10.

6. The cost of acid is 1.0¢ / 1,000 gal. Total chemical costs are 6.2¢ / 1,000 gal.

Comment: This final example, in comparison to example No. 3, shows the benefit derived from split chemical treatment for low phosphorus residuals.

7.5 References – Chapter 5

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Chapter 8

PHOSPHORUS REMOVAL BY LIME TREATMENT OF SECONDARY EFFLUENT

8.1 Description of Process

8.1.1 Theory

Lime treatment of wastewater is essentially the same process as the familiar lime softening of drinking water supplies. The objectives, however, are quite different. While softening may occur, the primary objective is to remove phosphorus by precipitation as hydroxyapatite. This reaction was described in Chapter 3.

During phosphorus precipitation other important reactions occur. The reaction of lime with alkalinity, that results in calcium removal when carrying out softening, not only takes place when treating wastewater, but may have a very important effect on the general efficiency of the process. This reaction can be considered to take place in the two following ways:

$$Ca(HCO_3)_2 + Ca(OH)_2 \rightarrow 2\ \underline{CaCO_3} + 2\ H_2O$$

$$NaHCO_3 + Ca(OH)_2 \rightarrow \underline{CaCO_3} + NaOH + H_2O$$

The first equation is that for softening. Some wastewaters do not contain enough calcium, however, for that equation to be satisfied. Calcium carbonate precipitation may still occur, but by the second reaction. The reactions to form $CaCO_3$ are important for two reasons: the lime consumption determines to a considerable extent the lime dose required for operating the process, and the resulting $CaCO_3$ acts as a weighting agent to aid in settling of sludge.

Another reaction that may be important is precipitation of $Mg(OH)_2$ as follows:

$$Mg^{2+} + 2\ OH^- \rightarrow \underline{Mg(OH)_2}$$

This reaction does not approach completion until the pH is raised to 11. Magnesium hydroxide is a gelatinous precipitate which aids colloid removal, but which hinders sludge thickening and dewatering. Where the pH must be raised to 11 or above to meet a phosphorus removal requirement or some other treatment objective, $Mg(OH)_2$ formation must be considered.

In the two-stage lime treatment process which will be described below, it is necessary to include recarbonation after the first stage to reduce the pH and precipitate the excess lime as $CaCO_3$ in the second stage. The following reaction occurs:

$$Ca^{2+} + CO_2 + 2\ OH^- \rightarrow \underline{CaCO_3} + H_2O$$

Carbon dioxide may also be used to lower pH after lime treatment. The important reaction in this case would be the conversion of Co_3^{2-} to HCO_3^-.

A final reaction that is of major concern, where lime is to be recovered by sludge recalcination, is as follows:

$$CaCO_3 \xrightarrow{\Delta} CaO + CO_2$$

The CaO produced would then be slaked to form $Ca(OH)_2$ before use.

8.1.2 Treatment Systems

Two lime treatment systems may be used with wastewater, single-stage and two-stage. Figure 8-1 shows a single-stage system. In single-stage treatment lime is mixed with feed water to raise the pH to a desired value. Although the pH will depend upon the required phosphorus removal, it is likely to be substantially less than 11, and may be less than 10. Precipitation of phosphate and other materials, as indicated by reactions discussed above, takes place. Time is allowed in the appropriate equipment for precipitate particles to flocculate to sufficient size for good settling. The clarified water from the settler may be discharged directly or may be filtered to improve solids removal. Adjustment of pH with CO_2 may be necessary before discharge and will almost certainly be required before filtration to prevent post-precipitation of $CaCO_3$ from the unstable water. The settled lime sludge may be disposed of as landfill or may be recalcined for recovery of lime. In the latter case, the sludge is thickened, dewatered by centrifuge or vacuum filter, and calcined. The calcined product is then slaked and reused. To avoid buildup of inerts in the lime, some of the sludge or recalcined lime must be wasted or the inerts must be separated from the sludge before calcination.

Two-stage treatment is somewhat more complicated than single-stage treatment. In typical two-stage treatment, shown in Figure 8-2, enough lime is added to the water in the first stage to raise the pH above 11. Precipitation of hydroxyapatite, $CaCO_3$, and $Mg(OH)_2$ occurs. Consideration of the solubility product for $CaCO_3$ and the equilibrium between CO_3^{2-} and HCO_3^- shows that the minimum solubility for Ca^{2+} occurs at a pH of about 10. At a pH of 11 or above there is a considerable concentration of Ca^{2+} present in the water. In two-stage treatment CO_2 is added after the first-stage settler to bring the pH down to about 10 where $CaCO_3$ precipitation results. The $CaCO_3$ is settled out and the clarified water is either discharged or sent to filtration. As in the case of single-stage treatment, pH reduction may be necessary before discharge and probably would be required before filtration.

8.2 Typical Performance Data

A number of full-scale and pilot-scale tertiary lime treatment plants are in operation and more full-scale plants are beginning operation. Only one plant, that at South Lake Tahoe, California, has operated for a significant period with recalcination of sludge for lime recovery. Most aspects of this 7.5 mgd plant have been discussed in detail by Culp and Culp (1). Although the data on handling of wastewater sludges for recalcination and the recalcination process itself are limited mainly to this plant,

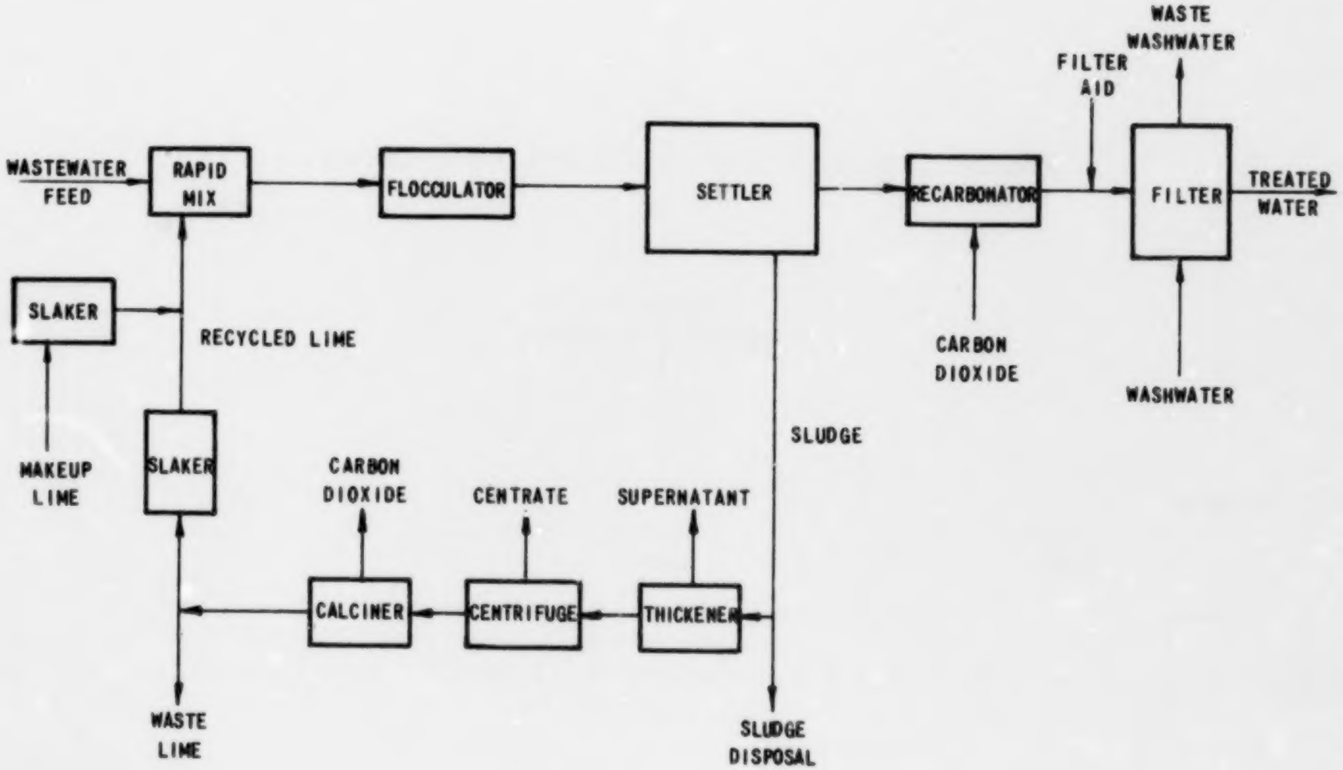
WASTEWATER FEED
RAPID MIX
FLOCCULATOR
SETTLER
RECARBONATOR
FILTER AID
WASTE WASHWATER
FILTER
TREATED WATER
CARBON DIOXIDE
WASHWATER
SLAKER
RECYCLED LIME
MAKEUP LIME
SLAKER
SLUDGE
CARBON DIOXIDE
CENTRATE
SUPERNATANT
CALCINER
CENTRIFUGE
THICKENER
WASTE LIME
SLUDGE DISPOSAL

FIGURE 8-1 SINGLE STAGE LIME TREATMENT SYSTEM

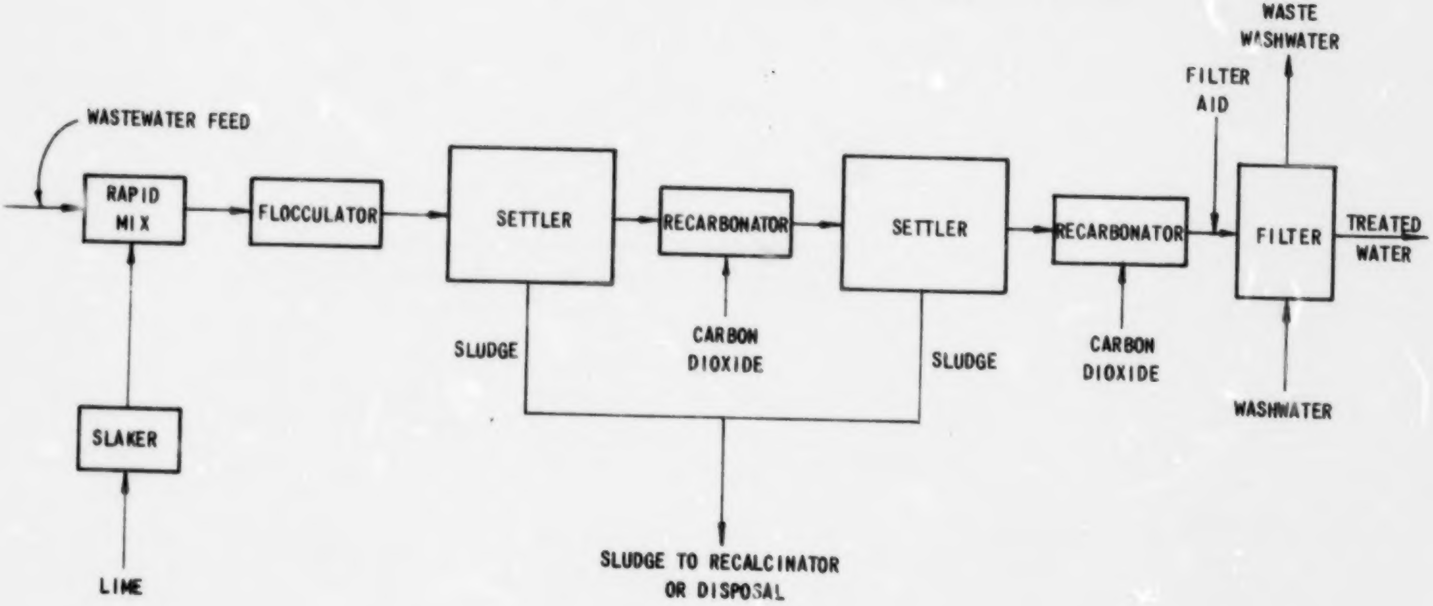
WASTEWATER FEED
RAPID MIX
FLOCCULATOR
SETTLER
RECARBONATOR
SETTLER
RECARBONATOR
FILTER
TREATED WATER
WASTE WASHWATER
FILTER AID
WASHWATER
CARBON DIOXIDE
CARBON DIOXIDE
SLUDGE
SLUDGE
SLUDGE TO RECALCINATOR OR DISPOSAL
SLAKER
LIME

FIGURE 8-2 TWO STAGE LIME TREATMENT SYSTEM

performance data on other parts of a lime treatment system are available from more than one location. Extensive data have been reported from pilot plants located in Washington, D. C., (2) and Lebanon, Ohio (3).

The lime treatment system at Tahoe is a two-stage clarification system which is followed by pressure filters of the multimedia type. These contain 3 ft. of a mixture of coal, sand, and garnet. The water passes through two filters in series. The first stage clarifier is operated at a pH of 11. To reach that pH requires about 300 mg/1 CaO. A small amount of polymer is used to improve flocculation. The pH is reduced to 9.6 by recarbonation before the second-stage settler. Before filtration, the pH is further reduced to 7.5. A small amount of alum (1 to 20 mg/1) or a combination of alum and polymer is used as filter aid. Filter run length has varied between 4 and 60 hours.

Phosphorus removal at this plant always has been good and has improved as operating experience has increased. By returning plant waste streams containing precipitated phosphorus to the first stage flocculator, it is now possible to obtain routinely an effluent with less than 0.1 mg/1 P. Before the filters the phosphorus concentration is about 0.4 mg/1.

In addition to phosphorus removal, there is significant removal of organic materials. During a period of intensive study of the first stage of clarification (unpublished data) 77% removal of BOD and 61% removal of COD were obtained. Although there was not a large removal of suspended solids, there was a significant change in the character of the solids from organic to largely inorganic. Further removal of organic materials and suspended solids occurs over the remainder of the system. Culp and Culp report typical filter effluents with a BOD of 3 mg/1, COD of 25 mg/1, and turbidity of 0.3 JTU.

Sludges from the first stage settler and the settler following recarbonation are sent to a gravity thickener. Solids concentration increases during thickening from about 1% to from 8 to 20%. Thickened sludge is then centrifuged to form a cake with from 30% to more than 40% solids. The cake is next calcined in a multiple hearth furnace and the recovered lime slaked for reuse. Initially it was planned to operate the centrifuge for high solids recovery in the cake. This necessitates discarding a significant amount of the recovered lime to prevent buildup of precipitated phosphate and other inerts. Since operation began, it has been found that much of the phosphate and some other inert materials can be separated from the $CaCO_3$ in the sludge by operating with a rather high solids concentration in the centrate. Approximately 90% of the phosphorus can be removed in the centrate, for example, when 25% of the solids entering the centrifuge are allowed to remain in that stream. This classification procedure results in a loss of about 15% of the recoverable lime. A greater degree of utilization of recalcined lime compensates for the loss, however, and the load on the calcining furnace is reduced. Over three years of operation, mostly without classification in the centrifuge, the average concentration of CaO in the recalcined product was 66%. Recalcined lime made up 72% of the total used at the plant.

Considerable operating data have been obtained by O'Farrell and Bishop (2) on a two-stage system operated at the EPAWQO-DC Pilot Plant in Washington, D.C. This 50,000 gpd plant treated secondary effluent from a pilot activated sludge system at the municipal treatment plant. The first stage pH was maintained at about 11.7, using a lime dose of about 400 mg/l as CaO, and the pH after recarbonation was maintained at about 10.3. Ferric iron was added at a rate of 5 mg/l in the second stage to improve flocculation. Water from the second-stage settler was filtered without further pH adjustment through gravity-flow dual-media filters consisting of anthracite coal over sand. The average filter run length was more than 50 hours.

Phosphorus removal for the system was similar to that obtained at Tahoe; 0.09 mg/l as P was the average concentration remaining. Phosphorus concentration before filtration averaged 0.13 mg/l.

In addition there was significant organic and suspended solids removal. The average BOD was reduced from 15 mg/l to 2.1 mg/l before and 1.5 mg/l after filtration. Suspended solids were reduced from 33 mg/l to 17 mg/l before filtration and 3.8 mg/l after filtration.

Sludge from the second stage settler was returned to the first stage and all sludge removed from the system was from the settler of the first stage. This sludge contained about 5% solids and constituted about 1.5% of the feed volume. Although there was sludge thickening and calcining equipment available, only preliminary study was made of lime reuse. Limited results showed improved thickening of the sludge when recalcined lime was recycled to the system.

A 75 gpm single-stage pilot system was operated at the Lebanon, Ohio Sewage Treatment Plant. Feed water was activated sludge effluent. This system, consisting of a flocculator-clarifier followed by dual-media filters of anthracite coal and sand, is described by Berg, et al. (3). Sludge was gravity thickened and discharged to sand drying beds. The system has been operated over a pH range from about 9 to 11 with most data taken at a pH of 9.5. Before filtration the pH was reduced to about 8.8 with sulfuric acid to prevent precipitation on the filter media. Filter run length averaged about 90 hours.

Effluent phosphorus concentrations for the system are shown in Figure 8-3. The effect of pH is clearly indicated. Since the average phosphorus concentration of the secondary effluent was about 10 mg/l as P, approximately 95% removal was obtained at a pH as low as 9.5. Before filtration the average phosphorus concentration was 0.75 mg/l at a pH of 9.5. Removal improved only slightly at higher pH because of the presence of precipitated phosphorus in the effluent.

Average suspended solids concentration of the secondary effluent was 43.5 mg/l. This was reduced to 16.5 mg/l in the clarifier. Much of the suspended matter in the clarifier effluent consisted of inorganic precipitates. After filtration the water was of high clarity. During a six month period when the biological treatment plant operated well, turbidity averaged 0.2 JTU. Even when the suspended solids load from the activated

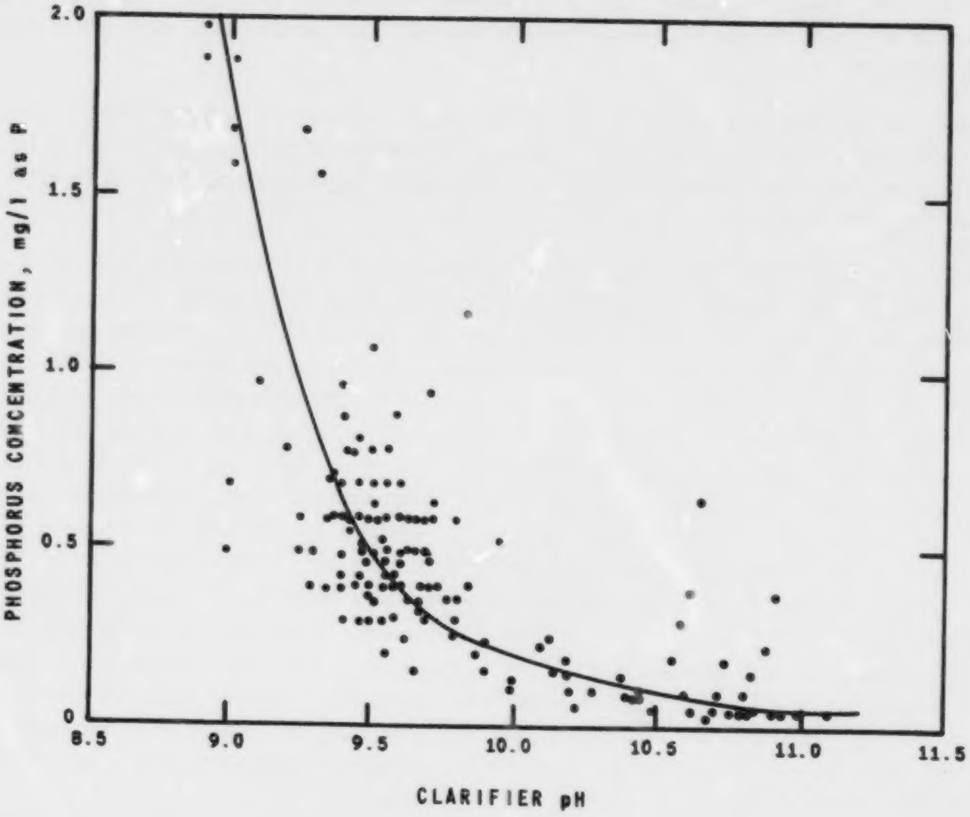

FIGURE 8-3 EFFECT OF pH
ON PHOSPHORUS CONCENTRATION
OF EFFLUENT FROM FILTERS
FOLLOWING LIME CLARIFIER

sludge plant was high, the lime treatment system produced water with a daily average turbidity not exceeding 2.5 JTU.

During a two month period in which extensive measurement of organic removal was made, total organic carbon was reduced by 63% over the clarifier and 68% over the clarifier and filters. Organic carbon measurements taken at other times indicated removals of from 55 to 74%. Occasional BOD and COD samples indicated removals of 86% and 62%.

Sludge was usually removed from the settler at 1.5 gpm giving a sludge concentration of 2.8% solids. This was thickened to about 10% before being pumped to drying beds. The sludge dewatered quickly to about 50% solids. Equipment was not available for recalcining the sludge.

The lime requirement is an important consideration in lime treatment. This can vary over a wide range depending on operating pH and water composition. From reactions given earlier it was seen that alkalinity has an important effect on the lime dose. Buzzell and Sawyer (4) have shown, for example, that for several wastewaters the lime dose required to reach a pH of 11 correlated approximately with alkalinity. Examination of the earlier equations would show also that calcium hardness can affect lime dose. One part by weight of CaO can react with from 0.89 to 1.79 parts of bicarbonate alkalinity expressed as $CaCO_3$, the lower value applying to very soft waters and the higher value to very hard waters. In addition to the reaction of lime with hardness, other competing reactions occur in lime treatment of wastewater. Also, there may be incomplete reaction of the lime. All of these complications make calculation of lime dose difficult. The result is that, at present, determination of lime dose is largely empirical. Approximate values have already been given for the plants at Tahoe and Washington, D.C. Some approximate values for the Lebanon work are shown in Table 8-1. It appears from the data already available that the lime dose will usually be in the range of 300 to 400 mg/l as CaO for two-stage treatment, and from 150 to 200 mg/l where single-stage treatment is satisfactory.

Table 8-1

LIME REQUIREMENTS

Feed Water Alkalinity (mg/l as $CaCO_3$)	Clarifier pH	Approximate Lime Dose (mg/l of CaO)
300	9.5	185
300	10.5	270
400	9.5	230
400	10.5	380

8.3 Criteria for Selection of Process

Lime treatment of secondary effluent represents a significantly higher capital cost and somewhat higher operating cost for phosphorus removal than mineral addition to a conventional treatment plant. Lime treatment has, however, several advantages over mineral addition. It can be considered more dependable since it adds additional flocculation and sedimentation steps to the system. Upsets in the conventional plant, which would reduce the efficiency of phosphorus removal by mineral addition, would have less effect on tertiary lime treatment. Because tertiary lime treatment is separate from the conventional treatment plant, it adds flexibility to operation of the system. For very high degrees of phosphorus removal, lime treatment would be the method of choice, not only because of the inherent greater dependability mentioned above, but because of the ability to produce an effluent slightly lower in phosphorus content. Lime treatment decreases the total dissolved solids content of the water by removal of hardness and alkalinity while mineral addition adds to the total dissolved solids. In the case of alum addition, for example, 5.3 parts by weight of sulfate are added for each part of aluminum. Lime treatment has the capability of removing turbidity to very low levels. Where there are plans for recreational reuse or certain industrial reuses of the treated water, low turbidity along with low phosphorus content of the lime treatment effluent is very desirable. Lime treatment offers the opportunity for recovery of the treatment chemical. At the present time there are no acceptable methods for recovery of aluminum or iron salts.

The choice of single-stage or two-stage lime treatment depends partly upon the degree of phosphorus removal required, but more importantly, on the alkalinity of the water. Unless a high treatment pH is used, waters with low alkalinity, in the range of 150 mg/l as $CaCO_3$ or less, form a poorly settleable floc because of the low fraction of dense $CaCO_3$. A pH above 10 is needed even to obtain measurable $CaCO_3$ production. A pH of 11 or above is then needed to precipitate $Mg(OH)_2$ which aids in settling of fine particles. Since neither discharge nor reuse of the high-pH water is likely to be acceptable, addition of a second treatment stage after recarbonation to a pH of about 10 becomes generally necessary. It would be possible to recarbonate to a much lower pH and avoid the second treatment stage, but this would result in a high calcium effluent and would eliminate the chance to produce high $CaCO_3$ sludge in situations where lime recovery was contemplated. Although most of the phosphorus removal occurs in the high pH first stage, in accordance with the pH dependence of phosphorus solubility as shown earlier by Figure 8-3, some removal does also occur in the second stage. Another important reason for including the second stage is to assure better control of clarification. With low alkalinity waters there is sometimes difficulty in settling the sludge in the first stage even at high pH. The second stage settler with its high $CaCO_3$ sludge, prevents solids carryover when the first stage settler does not operate properly.

With high alkalinity waters, a well settling floc is formed at pH values as low as 9.5. There is no need for two-stage treatment unless the required degree of phosphorus removal necessitates a high pH. In addition to obtaining a small degree of phosphorus removal, the second stage would be used in these cases to lower calcium content in the effluent and to obtain high $CaCO_3$ sludge for lime recovery. There is not yet available operating experience at enough plants to state positively the alkalinity at which single-stage treatment will perform satisfactorily, from the standpoint of floc settleability. At present, experience indicates that at an alkalinity of 150 mg/l as $CaCO_3$, single-stage treatment probably cannot be used. Between 150 mg/l and 200 mg/l settleability will probably depend on the amount of organic floc present. Above 200 mg/l, settleability is likely to be satisfactory.

8.4 Description of and Criteria for Choice of Equipment

8.4.1 Single-Stage System and First Stage of a Two-Stage System

Diagrams for single and two-stage lime treatment systems are shown in Figures 8-1 and 8-2. The first part of each system is made up of clarification equipment essentially the same as that found in water treatment plants. It includes a chemical feeding system (see Chapter 10), a rapid mix tank, a flocculator, a settler, and a means for sludge removal. Mixing, flocculation, and settling may be carried out in separate vessels or tanks, or they may be combined in one integrated unit. The system consisting of separate tanks has been used for many years in water treatment and was the system adopted for the first stage of the Tahoe lime treatment plant. The integrated type of equipment is sometimes referred to as an upflow clarifier or a sludge blanket clarifier. The latter term can be misleading, however, since not all such units operate with a sludge blanket. A diagram of a unit that does not operate with a sludge blanket is shown in Figure 8-4. Culp and Culp (1) recommend the system of separate tanks because that system allows separate control of each part of the process. They point out that such a system has greater flexibility in points of addition for chemicals. They also point out that in the integrated units with a sludge blanket there may be difficulty in controlling the blanket height and the blanket may become anaerobic, leading to poor phosphorus and solids removal. Pilot studies at other locations have shown that excellent solids removal can be obtained with sludge blanket equipment, but that blanket instability could be a problem. On the other hand, pilot upflow units designed to operate without a sludge blanket have proved to be very effective at Washington, D. C. (2). Results available at this time indicate that both the system of separate tanks and the integrated units without sludge blankets should be considered for design of new plants.

The Tahoe plant is an excellent example of a system with the first stage consisting of separate tanks for each unit process. At the present time the plant must be relied upon heavily as a guide to design of such systems. Description of the equipment and some design parameters are given by Culp and Culp. The rapid mix tank which is located in one corner of the flocculator has about a 30 second residence time at design flow. (The plant usually runs with a daily average flow of about one third of design flow.) Mixing is accomplished with a vertical shaft mixer. The flocculator is a

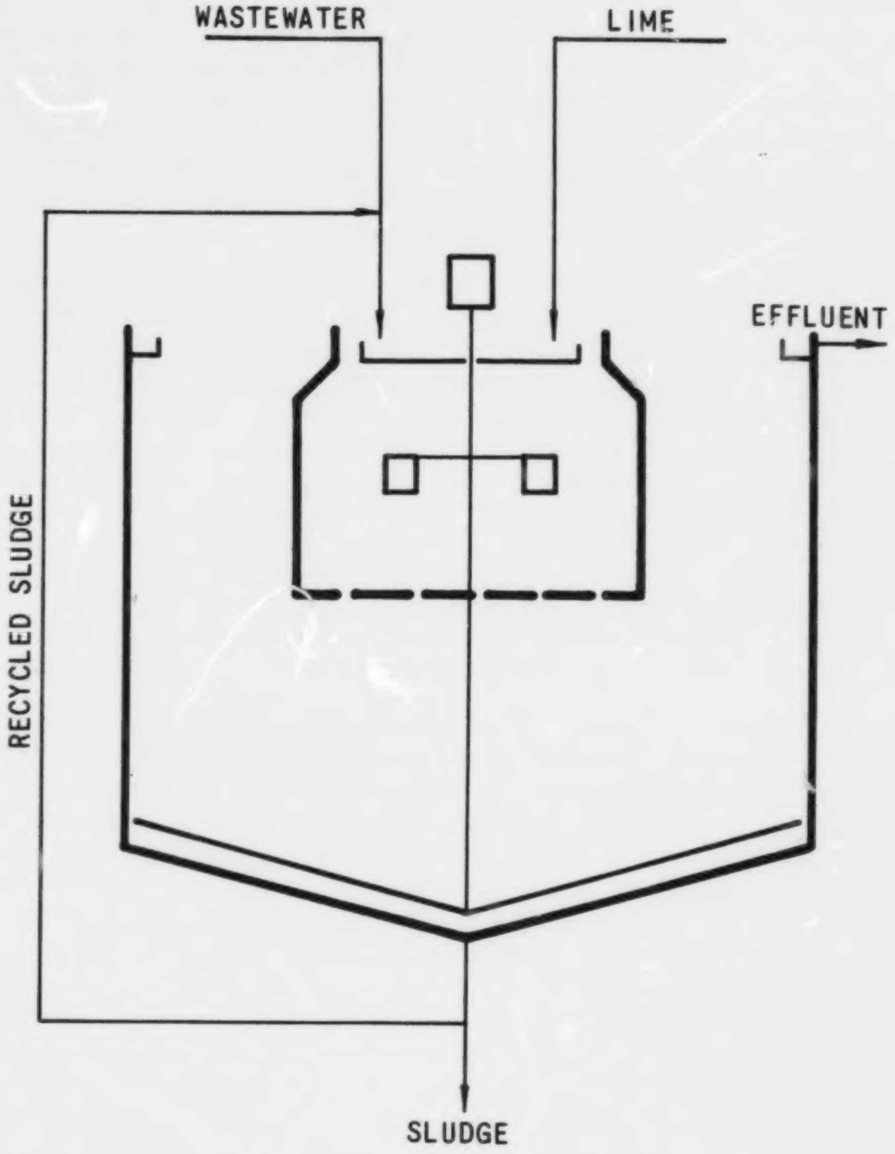
WASTEWATER
LIME
EFFLUENT
RECYCLED SLUDGE
SLUDGE

FIGURE 8-4 TYPICAL UPFLOW CLARIFIER

square tank with a depth of 8 ft. At design flow the residence time is 4.5 minutes. The flocculator is provided with air agitation, but operating experience has shown that this agitation is not necessary. The settler is circular with a center inlet. The depth is 10 ft and the design overflow rate is 950 gpd/ft^2. There are two sludge pumps, one a variable-speed centrifugal, and the other a positive displacement Moyno. Provision has been made to return part of the sludge to the rapid mix tank. The concept of returning sludge to the section of the clarification equipment where precipitation is taking place is called solids contact. The objective is to hasten precipitation and to obtain larger precipitate particles which will settle well.

The ease with which flocculation occurs at Tahoe would suggest that a flocculator for this service presents no design problems. Experience is, however, limited. Pilot testing would be very desirable to determine the flocculating characteristics of a particular wastewater. Jar tests may be of some value. Unfortunately, the effect of solids contact, which experience at Tahoe and elsewhere indicates is beneficial, is difficult to duplicate in a jar test. If jar tests indicated good flocculating characteristics without benefit of solids contact, however, flocculation should be as good or better with solids contact. When pilot studies are carried out, provision should be made for solids contact.

Comments concerning the effect of solids contact in jar tests also apply to settling rates determined from these tests. Rapid settling without benefit of solids contact would be a strong indication that settling with solids contact would also be rapid.

Clarification equipment of the integrated type for use in municipal water treatment is manufactured by a large number of companies. Units of this type used for treatment of wastewater have been essentially of the same design as used for municipal water supplies. The basins are usually circular with the rapid mix and flocculating sections located at the center of the settler. Agitation for mixing and flocculating as well as power for sludge scrapers is provided at the center of the settler. Solids contact is usually provided, and may be carried out with best control by external circulation of sludge to the mixing section. In some designs, however, internal recirculation is used. There is considerable variation in the geometry and complexity of the mixing and flocculating sections depending upon the manufacturer. The unit shown in Figure 8-4 is a relatively simply type. Units from three different manufacturers were used in pilot work at Lebanon, Ohio and Washington, D.C. These and others at additional locations have performed effectively. However, in some cases with operation of a sludge blanket, difficulties have been encountered. A sludge blanket can be eliminated by avoiding designs in which the wall separating the center compartments from the concentric settler reaches close to the bottom of the settler. The design in Figure 8-4 prevents sludge blanket formation.

There has been little effort to modify this type of equipment from use in water treatment to use with wastewater. Experience at Tahoe would suggest, for example, that less flocculation time, and probably less agitation, is required than is generally provided. A unit such as that shown by Figure 8-4 may have about 40 minutes flocculation time at design capacity for water treatment, nearly nine times that at

Tahoe. Whether modifications in such equipment would be worthwhile remains to be seen.

If a designer chooses to use equipment that is on the market, he does not have complete flexibility in specifying the design. Fortunately, equipment satisfactory for water supply treatment has proved satisfactory for wastewater, except that a lower settler overflow rate is necessary. The single-stage treatment system at Lebanon, Ohio was operated at a constant overflow rate of 1,440 gpd/ft^2. This proved satisfactory even with a sludge blanket At Washington, D.C., tests were made with settler rates as high as 1,950 gpd/ft^2 although the average rate was 1,120 gpd/ft^2. For design, a peak, dry weather overflow rate of 1,200 to 1,400 gpd/ft^2, with the average somewhat lower, appears reasonable.

8.4.2 Recarbonation

In a two-stage system recarbonation follows the first stage settler. In a single-stage system and following the second stage of a two-stage system, pH reduction by recarbonation may be necessary to make the effluent suitable for filtering or for discharge. An excellent discussion of all aspects of wastewater recarbonation is given by Culp and Culp (1). The reader is referred to that publication, as well as one by Haney and Hamann (5), especially for information on sources of carbon dioxide. Sources include stack gas from sludge incinerators and lime recalciners, CO_2 generators, and commercial liquid CO_2.

The equipment used for contacting the CO_2 and water may be simply a tank with the gas being bubbled through the water. This is the system used at the Tahoe plant where the height of water over the CO_2 source is 8 ft. Pilot studies at Washington, D.C., with a tank depth of 11 ft and a turbine mixer to reduce bubble size and distribute bubbles, showed almost 100% absorption of CO_2.

When considering residence time, the recarbonation tank following the first stage of a two-stage system must be differentiated from the recarbonation tank used just to reduce effluent pH. The residence time of the recarbonator between the first and second stages is not particularly important. At Tahoe the recarbonation tank is only large enough to give a 5 minute residence time at design flow. In the Washington, D.C. studies, the residence time was 15 minutes. A problem that may arise when recarbonating wastewater is foam formation. This is most pronounced when the flow of bubbles is concentrated in parts of the recarbonation tank. The CO_2 distributing system should cover as well as possible the whole lateral area of the tank bottom. This would be especially true if the residence time were as low as 5 minutes.

In the recarbonation tank used just for effluent pH adjustment, sufficient residence time must be allowed for completion of reactions taking place. Culp and Culp recommend 15 minutes. In the Tahoe plant only 4 minutes is provided at design flow, but the water flows to storage ponds with a much longer residence time before the water is filtered. Just as in the case of recarbonation between stages of a two-stage system, good bubble distribution is important for prevention of excessive foaming.

Since one reason for using a two-stage system is to prevent a high calcium concentration in the effluent, the pH to which the water should be recarbonated between stages is that for maximum conversion of the calcium to $CaCO_3$. Where lime recovery is practiced, maximum formation of $CaCO_3$ is also desirable because it results in maximum CaO production. The optimum pH is about 10. At Tahoe, for example, 10.3 was selected because tests showed that pH to give maximum $CaCO_3$ production.

The pH to be selected when the objective of recarbonation is just pH reduction, will depend on a number of factors. An effluent standard may determine this pH. If stabilizing the water to prevent post precipitation is the objective, the pH may be determined by consideration of the Langelier Index. If, as in the case of Tahoe, a treatment step such as activated carbon adsorption is to follow lime treatment, this will determine pH. In the latter case a pH of about 7.5 has usually been selected.

The CO_2 dose requirements can be calculated from the chemical reactions taking place and a knowledge of the concentrations of the various forms of alkalinity in the water. For design of a new plant, alkalinity data must be obtained from samples of liquor taken from jar tests run at the same pH values as planned for the plant. Results must be considered very approximate, but they do help in sizing CO_2 feeding equipment. Culp and Culp discuss the calculations in detail. It was found in pilot plant work at Washington, D.C. that the CO_2 dose for recarbonation between the stages of a two-stage treatment system can be calculated reasonably well by considering the reduction in calcium concentration that occurs. This reduction is due to $CaCO_3$ formation with the CO_3^{2-} coming from the CO_2. The calcium content before recarbonation can be determined satisfactorily from jar tests run at the desired operating pH. The calcium content after recarbonation can also be determined approximately by jar test. Results from several pilot plants indicate the soluble calcium content at the pH of minimum solubility should be about 40 mg/l as Ca. This figure is probably just as good for calculations as a figure obtained from a jar test. The CO_2 dose in mg/l is then equal to:

$$\text{(Ca reduction in mg/l)} \left(\frac{44}{40}\right) .$$

A safety factor of about 20% should be added to the calculated dose to compensate for inefficiency in absorption.

8.4.3 Second Stage Flocculation and Sedimentation

In two-stage treatment, recarbonation is followed by the second stage flocculation and settling. Experience with equipment for the second stage of treatment with wastewater is limited. Two very different systems have been tested. The simplest is the system at Tahoe. In the Tahoe plant the equipment used is just a longitudinal settler with a 30 minute detention time and a 2,400 gpd/ft^2 overflow rate at design flow. No flocculation equipment is provided. Culp and Culp (1) refer to the settler as a reaction and settling basin. It has been found at Tahoe that a significant part of the recarbonation reaction actually occurs in the settler. This is shown by the pH decrease of about 0.7 unit that occurs between the recarbonation tank and the settler outlet. The clarification that results at Tahoe is unusually good for the simplicity of the equipment.

Jar test and pilot plant work at Washington, D. C. indicated that flocculation after addition of a flocculating aid was required to obtain satisfactory clarification of that water. The pilot tests were run in upflow equipment of the integrated type discussed earlier. Without a flocculating aid, fine $CaCO_3$ precipitate escaped the settler, even at a settler overflow rate as low as 1,100 gpd/ft^2. Using 5 mg/l of Fe^{3+}, good operation was observed at overflow rates as high as 1,950 gpd/ft^2. Hydraulic limitations prevented testing at higher rates. Apparently, the good contact between solids and water during flocculation was beneficial in reducing $CaCO_3$ supersaturation in the effluent. Precipitation in the filters following the second stage did not occur, and no pH reduction was required.

In this work the sludge from the second stage was returned to the mixing section of the first stage to serve as a weighting agent. In waters of low alkalinity or expected high loads of biological solids, this procedure should be considered.

It is difficult to recommend the amount of flocculation to provide and the overflow rate to use for second-stage treatment. Jar tests would be of little value because of the complication of first raising the pH with lime and then lowering with CO_2. Pilot testing may be desirable. Such a pilot system would have to include the first stage equipment and recarbonation equipment, although these could be of crude design. In the absence of pilot tests to indicate otherwise, flocculation should be provided in the second-stage system. If filtration is included, a peak dry weather overflow rate of up to 2,000 gpd/ft^2 in the settler may be used. If filtration is not provided, a somewhat lower overflow rate should be chosen to assure reasonable effluent quality during periods when an upstream part of the system is not operating properly.

8.4.4 Filtration

The last step in a complete lime treatment system is filtration. Although there are a variety of filter types that could be used, essentially all the recent test work has been done with downflow filters of multimedia type. These are essentially the same in design as a rapid sand filter, except that the media are graded from coarse at the filter surface to fine at the filter outlet. This is accomplished by using media of different densities with the largest particles being composed of the least dense material. The result is a filter bed which has far more capacity for removing suspended solids than an ordinary sand filter, without impairing effluent clarity. Both gravity and pressure filters have been used. Each has certain advantages. Generally, however, pressure filters are more appropriate at smaller plants.

Two media combinations have been tested extensively. These are dual-media of coal and sand, and tri-media of coal, sand, and garnet. Filters used at Lebanon, Ohio following a single-stage system (3) and filters at Washington, D. C. following a two-stage system (2) were of the dual-media type. Both contained 18 in. of anthracite coal over 6 in. of sand. Average particle sizes of the media at Lebanon were 0.75 mm and 0.46 mm; at Washington, D.C., 0.9 mm and 0.45 mm. At 2 gpm/ft^2 the average run length at Lebanon was about 90 hours using 8 ft of water as the terminating pressure loss. At Washington, D.C., filter run length averaged about 50 hours for an average rate of 3.4 gpm/ft^2 and a terminating pressure loss of 7 ft of water. High clarity waters were obtained in each case. Backwash in both cases was carried out at 20 gpm/ft^2. Backwash time was 5 minutes at Lebanon and 10 minutes at Washington, D. C. In the latter case, a surface wash was included.

These run length figures give a rough idea of the results that may be expected at other locations. For maximum run length at a given product quality, the depths of media and their particle sizes should be optimized. Although this was not done for the above-mentioned filters, pressure loss distribution in the filters at Lebanon as reported by Berg, et al. (3) indicates that good solids storage in the anthracite was being obtained. Small pilot filters could be used for optimization of the media.

A reasonable design rate for gravity-flow dual-media filters appears to be about 3 gpm/ft^2. Higher rates could result in inconveniently short runs during periods of high solids load. Rates much lower than 3 gpm/ft^2 result in excessive filter costs, not justified by the longer filter runs. Very long filter runs could result in backwash problems from biological activity in the filters. Recommendation of a 3 gpm/ft^2 rate is based upon having good operation of the second-stage settler. Frequent settler upsets must be avoided.

Tri-media filters are used at the Tahoe plant. These are pressure filters which at design flow operate at 5 gpm/ft^2. Each bed holds 3 ft of mixed coal, sand, and garnet as supplied by Neptune Microfloc. Two filters are used in series, and are backwashed in series, usually when the head loss reaches 16 ft of water. Run length has varied from 4 hours during heavy solids load to about 60 hours under good conditions. Backwash is carried out at 15 gpm/ft^2. The backwash water is reprocessed through the lime treatment system. More details are given by Culp and Culp (1).

From the available data, it is difficult to give precise criteria for choosing between dual-media or tri-media filters following lime treatment. The use of garnet allows for a smaller particle size at the bottom of the filter than is possible with sand. In case of floc weakness, the tri-media filter offers, therefore, more protection from turbidity breakthrough. The cost, however, is slightly higher. Where the tri-media fill is not used, filters can still be backwashed at the onset of turbidity breakthrough. Experience with dual-media filters on lime treated water has not shown sudden breakthrough to be an important problem. Where water of the highest clarity is required, tri-media filters may be of most value.

8.4.5 Sludge Handling

The sludge from a lime treatment system may be handled in two general ways. It may be thickened, dewatered, and disposed of or it may be thickened, dewatered, and recalcined to recover lime for reuse. For small plants, recalcination will not be economically competitive with disposal and should not be considered unless there are restrictions on disposal which make that alternative difficult. It can be assumed generally that for plants over 10 mgd, recalcination of lime sludge will be practical. Recalcination may also be practical at plants somewhat smaller than 10 mgd, depending on the cost of purchased lime and other local conditions.

Data given earlier indicate that the volume of sludge from lime treatment will vary from 1.5% to several percent of feed volume. Sludge concentration will probably be in the range of 1% to 5%. The actual weight of sludge will vary with the chemical composition and amount of suspended solids in the feed water. Values have been observed in the range of 4 to 7 lb/1,000 gal. Sludge production from the first stage of a two-stage system or from a single-stage system can be estimated approximately by weighing the dried sludge from jar tests run at the planned operating pH. Additional sludge produced in the second stage can be assumed to be the $CaCO_3$ formed from calcium concentration reduction during recarbonation. Calcium content before recarbonation can be obtained from the above-mentioned jar tests and after recarbonation can be assumed to be 40 mg/l as Ca.

Design information for the Tahoe sludge handling system is reported by Culp and Culp. (1) The reader is referred to that publication for further information. The Tahoe plant is the only one which includes recalcination of sludge from the lime treatment of wastewater, and that has been operated for a long period of time. Paramenters from that system can be used as a rough guide for design. The gravity thickener has a bottom scraper mechanism and is 8 ft deep. Overflow from the thickener is returned to either the primary clarifier or the first stage of lime treatment. The design solids loading is 200 lb/day/ft^2 and the design overflow rate is 1,000 gpd/ft^2. Some preliminary data from pilot work at Washington, D.C. suggest these loadings are high. The Tahoe equipment was sized, however, to handle a volume of sludge equal to about 9% of the plant design flow. The actual volume of sludge should usually be less than one third of that volume. At Washington, D.C., recalcined lime was found to produce a sludge which thickened significantly faster than sludge from virgin lime.

If lime is not to be recovered from the sludge, the underflow from the thickener can be placed directly on drying beds for final dewatering, or it can be dewatered by centrifuge or vacuum filter.

If sludge recalcination is planned, the thickened sludge would be dewatered, by centrifuging or filtering and fed to the calcining furnace. The centrifuge may have advantages over filters for this purpose, such as the ability to separate phosphate sludge from $CaCO_3$. Lime from the furnace is stored for reuse. It, along with any new lime, must be slaked before use. Sludge can be pumped from the thickener to the centrifuge. Cake from the centrifuge must be transported by conveyor. At Tahoe the centrate is sent to the primary clarifiers. It has been found at Tahoe that, by operating the centrifuge with less than maximum solids capture, much of the precipitated phosphorus can be retained in the centrate. This results in a higher quality lime. This mode of operation requires a second centrifuge to remove the phosphorus rich solids from the centrate. These solids may have value for use in fertilizer.

The Tahoe dewatering and recalcining system is described by Culp and Culp. The centrifuge is a 24 by 60 in. concurrent flow type. The calciner is a 14 ft-3 in. diameter, 6 hearth furnace operated at a top temperature of 1,850° F. Other types of furnaces have been used in water treatment plants and these should also be applicable to the sludge from wastewater treatment. The reader is referred to equipment manufacturers for further information about centrifuges and furnaces.

8.4.6 Control of Lime and Carbon Dioxide Feed

The feeding of lime to a lime softening system can be controlled in a number of ways. For the wastewater application, the most appropriate appears to be control of pH, with the pH value being selected for good suspended solids removal or to meet a phosphorus removal requirement. Where flow equalization is employed, pH sensing alone should be sufficient. Where there is substantial diurnal variation in flow, better control can be maintained by flow proportional-pH control. Systems are available for this type of control.

Carbon dioxide dose is also controlled by pH. As in the case of lime feed, flow proportional-pH control would be preferred where there is diurnal variation in flow.

There is a tendency for pH electrodes to become coated with precipitates and lose sensitivity. The precipitate can be dissolved with acid. In some instances, however, it has been found satisfactory to use manual control because of difficulties with the pH control system.

8.4.7 Scale Formation on Equipment

Because the water in a lime treatment system is supersaturated to some degree with $CaCO_3$ and other precipitating substances, there is a tendency for scale to form on equipment and pipe surfaces. The problem is particularly serious with the lime slurry from the slaker. This can quickly plug the slurry line to the rapid mix tank. There should be easy access to all parts of this line for cleaning. At one small plant a flexible hose was used to feed lime slurry. By periodically flattening or flexing the hose, the scale was removed. The mechanical mixer in the rapid mix tank will also become scaled and must be cleaned.

Scale can also form in sludge lines and the effluent line from the first-stage clarifier. It is recommended that open troughs be used wherever possible. Provision should be made for cleaning lines when open conduits are not possible.

Possible scaling of filters has already been mentioned. Recarbonation before the filters will minimize scale formation.

8.5 Capital and Operating Costs

Because of the short history of lime treatment of wastewater, there is a scarcity of capital and operating cost information. Costs are available for the Tahoe plant and are reported by Culp and Culp (1). Capital cost for the 7.5 mgd lime treating facility was $1,115,000 and cost of the filters was an additional $705,000. For 1969 the estimated operating costs exclusive of equipment amortization were 7.3¢/1,000 gal. for lime treatment without filtration and 2.8¢/1,000 gal. for filtration. Amortization of equipment, based on costs adjusted to the 1969 national average, interest of 5% for 25 years, and the assumption of the plant operating at full capacity, would add 2.7¢/1,000 gal. for treatment without filtration and 1.8¢/1,000 gal. for filtration. For the plant operating at full capacity the total cost of operation would then be 10.0¢/1,000 gal. without filtration and 14.6¢/1,000 gal. including filtration. The fraction of the cost resulting from amortization is a substantial 31% even with the assumption of full capacity. There is an obvious need to keep equipment size to a minimum. Flow equalization deserves strong consideration when planning for lime treatment of secondary effluent to minimize the initial cost of new equipment.

Additional costs for lime treatment based on information from Tahoe and other sources have been reported by Smith and McMichael (6). Tables 8-2 and 8-3 show costs based largely on their results. Capital costs have been updated to December, 1970. Amortization is at 6% for 25 years. For 1 mgd plants, recalcination equipment is

Table 8-2

CAPITAL COST OF LIME TREATING FACILITIES

Treatment	Cost ($) Plant Size (mgd) 1.0	 10	 100
Single-Stage without Filtration	100,000	1,200,000	5,500,000
Two-Stage without Filtration	160,000	1,500,000	7,900,000
Dual-Media Filtration	110,000	510,000	2,300,000

Table 8-3

TOTAL COST FOR LIME TREATMENT OF WASTEWATER

Treatment	Cost (¢/1,000 gal.) Plant Size (mgd) 1.0	 10	 100
Single-Stage without Filtration	13	7	4
Two-Stage without Filtration	16	9	6
Dual-Media Filtration	8	3	1.4

not included. All lime would be purchased. The applicability of recalcination was discussed earlier. Component costs making up the total cost figures represent an average for the whole country. Local conditions may cause significant deviation from these values.

8.6 References – Chapter 8

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Chapter 9

PHOSPHORUS REMOVAL BY MINERAL ADDITION TO SECONDARY EFFLUENT

9.1 Description of Process

Alum and to a lesser extent iron salts have been evaluated as precipitants for phosphorus in effluents from conventional treatment plants. Both batch studies and continuous-flow systems up to 2.5 mgd have successfully reduced phosphorus to very low levels.

Equipment typically consists of mixers, flocculators, settlers, and filters. Some work has been done, however, with the chemicals added directly to filters. Specific studies are discussed later and provide additional information on equipment commonly used.

9.2 Summary of Design Information

An alum dosage of about 200 mg/l is required for phosphate removal from typical municipal wastewater while dosages of 50 to 100 mg/l are sufficient for effluent clarification. Iron salts, while successful in precipitating phosphate, have found little application because of residual iron remaining in the treated water. An Al:P molar ratio of 1:1 to 2:1 is required. The optimum pH for alum treatment is near 6.0 while for iron the optimum pH is near 5.0 (1). The pH of high alkalinity waters may be reduced either by using high dosages of alum or by adding supplementary dosages of sulfuric acid. Anionic polyelectrolytes are useful with alum in improving settleability of floc. Settling alone will reduce residual P to about 1 mg/l.

If higher removals are required, filtration must be employed. With proper operation, residual P may be reduced to less than 0.1 mg/l. Multimedia filters are preferred over sand filters because of the extended length of run. Microstrainers have not been successful in removing alum floc.

Surface overflow rates for clarifiers have ranged from 580 to 1,440 gpd/ft^2, but for consistent removal of phosphate, the lower end of the range is preferred. Filtration rates of 2 to 5 gpm/ft^2 are used.

No process has yet been developed for successfully recovering alum from sludge for reuse although both alkaline and acid regeneration were investigated (2, 3). The sludge can be dewatered by conventional methods. Because dewatering may be difficult, design of such facilities should be conservative.

Costs for tertiary treatment for phosphate removal range from 6 to 15¢/1,000 gal. depending on plant size and degree of treatment required.

9.3 Laboratory and Pilot Studies

9.3.1 South Lake Tahoe (3, 4, 5, 6)

Liquid alum was used to precipitate phosphate in a 2.5 mgd tertiary plant at South Lake Tahoe, California. The treated wastewater was applied to multimedia filters for solids removal without prior sedimentation.

Chemical feed was proportioned automatically to flow, and flash mixing was provided by an in-line mixer in the filter influent line. The filter media consisted of layers of anthracite coal, graphite, fine garnet and coarse garnet overlaying a supporting bed of gravel. Two filters were operated in series at 5 gpm/ft^2. Backwash with secondary plant effluent was initiated either by headloss or by a rise in turbidity of the effluent. Polyelectrolyte was used as a filter aid to control the depth of penetration of floc into the filter beds. With an alum dosage of 200 mg/l, 8 to 9 mg/l P in the secondary plant effluent was reduced to 0.03 to 0.3 mg/l P in the filter effluent. Turbidities were correspondingly reduced from 30 to 70 JTU to 0.2 to 3.0 JTU.

Efforts to regenerate alum from the sludge by both acid and alkaline treatment proved to be uneconomical and a lime treatment process was developed as a substitute for the alum process. Lime could be economically recovered for reuse and had the added advantage of producing a high pH which facilitated removal of nitrogen by ammonia stripping.

Total costs for phosphate removal only, as estimated in 1966, are shown in Table 9-1.

Table 9-1

1966 ESTIMATED COSTS FOR TERTIARY
PHOSPHATE REMOVAL WITH ALUM

Plant Capacity (mgd)	Total Cost (¢/1,000 gal.)
2.5	11.7
10	8.5
50	7.1
100	6.6
200	6.3

9.3.2 Nassau County (7, 8)

Nassau County, New York, is evaluating treatment methods for its wastewater to provide a product water which is suitable for ground injection as a barrier to salt water intrusion. A pilot plant has been operated at 400 gpm producing a water which in most respects exceeds USPHS drinking water standards. Effluent from high rate activated sludge is pumped into a 40 ft diameter by 14 ft deep clarifier where alum and polyelectrolyte are added. Sludge is recirculated into the coagulation zone to improve efficiency of floc formation. The treated wastewater then passes downward through a flocculation zone and upward through a clarification zone to orifice type collectors. The clarified wastewater passes to 2 dual-media filters in parallel operated at 3.0 gpm/ft^2. Each filter consists of a 36 in. bed of No. 1-1/2 anthracite (effective size of 0.90 mm) above a 12 in. layer of sand (effective size of 0.40 mm). Filter backwash includes air scour, surface wash and high and low-rate backwashing. The filter effluent passes through 4 granular carbon contactors operating in series to remove organics.

Typical operating results with about 200 mg/l alum are shown in Table 9-2.

Table 9-2

NASSAU COUNTY PERFORMANCE DATA

	Influent	Effluent
Turbidity, JTU	20 - 25	0.2 - 1.5
Color, Pt-Co Units	20 - 40	none
COD, mg/l	78	13
P, mg/l	9	0.1 - 1.0

Costs have been estimated, for separate tertiary treatment only, at plant sizes of 1, 10, and 100 mgd as shown in Table 9-3.

Table 9-3

NASSAU COUNTY COST ESTIMATES

	Plant Capacity (mgd)		
	(1)	(10)	(100)
	(Costs, ¢/1,000 gal)		
Process			
Coagulation	4.9	3.5	3.2
Filtration	1.8	1.1	1.0
Carbon adsorption	6.3	4.5	4.0
	13.0	9.1	8.2
Operating labor	28.0	5.6	1.8
Total Cost	41.0	14.7	10.0

9.3.3 Chicago, Hanover Park (9)

Alum was used to coagulate solids and precipitate phosphate from secondary effluent at Chicago's Hanover Park tertiary treatment plant. Treatment consisted of coagulation-sedimentation in circular clarifiers followed by either rapid sand filtration or microstraining for polishing. Average annual flow was 1.5 mgd.

Combined coagulation plus sand filtration removed 34, 58, and 82% P at dosages of 38, 60, and 140 mg/l of alum, respectively. Combined coagulation plus microstraining effected a lower phosphate removal of 23% at an alum dosage of 38 mg/l. In general, alum coagulation under the test conditions gave little or no improvement in suspended solids removal over that obtained by sand filtration of uncoagulated secondary effluent. The microstrainer did not significantly remove alum-coagulated solids.

9.3.4 Dallas (10)

Dallas, Texas, is operating a demonstration project designed to provide necessary design data for upgrading of the Dallas-White Rock trickling filter plant. Phosphorus removal facilities include a solids contact upflow clarifier unit and two multimedia filters. The clarifier has a rise rate of 0.5 gpm/ft^2 at the design flow of 100 gpm and a detention time of 4 hours. Both filters contain anthracite over sand media. One is a standard gravity filter. The other is a biflow type, with feed water entering at both the top and bottom of the bed. Filtered water is removed by a manifold-lateral collector located at the mid-depth of the bed. The gravity filter is designed for 50 gpm at a filtration rate of 4 gpm/ft^2. The biflow filter is designed for 100 gpm at a filtration rate of 8 gpm/ft^2, or 4 gpm/ft^2 for both the bottom and top of the bed.

The pilot plant process has not yet been optimized for phosphorus removal, however, limited data from two short test periods on trickling filter and activated sludge effluents are shown in Table 9-4.

Table 9-4

REMOVAL OF TOTAL ORTHOPHOSPHATE BY COAGULATION-CLARIFICATION IN THE DALLAS DEMONSTRATION PLANT

	Trickling Filter Effluent	Activated Sludge Effluent
Lime Dose, mg/l CaO	128	133
$FeCl_3$ Dose, mg/l Fe^{3+}	25	56
Soluble Phosphate In, mg/l P	6.5	7.0
Soluble Phosphate Out, mg/l P	0.43	0.21
% Soluble P Removed	93.5	97.0
Test Period, days	12	2

9.3.5 Dayton (11)

Dayton, Ohio, used alum in a 310 gpm pilot plant to remove phosphate and suspended solids from trickling filter effluent. The rectangular sedimentation basin was designed for 105 minutes detention and an overflow rate of 1,014 gpd/ft^2. Two filters designed for loadings up to 8 gpm/ft^2 were used for polishing effluent. One contained 30 in. of sand while the other used 20 in. of anthracite overlaying 10 in. of sand. The sand had an effective size of 0.40 to 0.45 mm with a uniformity coefficient of 1.35 to 1.70. The anthracite had an effective size of 0.85 to 1.20 mm with a uniformity coefficient of 1.35 to 1.80. Both air and water were used in backwashing.

Alum with activated silica or anionic polyelectrolyte as coagulant aids achieved phosphate removals at a dosage rate of 5 lb alum/lb PO_4^{3-} down to residuals of 4 mg/l PO_4^{3-} in filter effluents (89% removal of phosphate). Further removals down to 1 to 2 mg/l PO_4^{3-} required in excess of 200 mg/l alum. The clarifier was operated at overflow rates of 655 to 1,035 gpd/ft^2.

The filters removed 30 to 70% of the applied phosphate load. The dual media filter produced an effluent comparable to that from the sand filter with filter runs 2 to 3 times the length of the sand filter runs.

Costs to provide 90% phosphate removal with alum for the present 50 mgd flow in a 75 mgd plant were estimated at 6.5 to 8.2¢/1,000 gal.

9.3.6 Oxidation Pond Effluents

At Lancaster, California, a research project was initiated to renovate wastewater for use in recreational lakes (12). Existing treatment consisted of primary sedimentation followed by 45 to 60 days detention in oxidation ponds. For coagulation, laboratory jar tests showed alum to be more effective on a cost basis than either lime or ferric sulfate. Phosphate residuals were reduced to 1.1, 0.2, and 0.01 mg/l P with alum dosages of 150, 200, and 300 mg/l, respectively, at an optimum pH near 6.0. Pilot studies showed that both sedimentation and air flotation would provide satisfactory separation of most of the precipitated phosphates, however, filtration was necessary to obtain the required low residuals of less than 0.15 mg/l P.

Final specifications for a full scale process included coagulation with about 300 mg/l alum and 20 minutes of flocculation with a conventional paddle flocculator, sedimentation in a horizontal clarifier with an overflow rate of 400 gpd/ft^2, and filtration through a dual media gravity filter consisting of 18 in. of 0.5 mm anthracite over 8 in. of No. 20 sand.

Capital costs of a 0.5 mgd facility were estimated at \$150,000 with operating costs of \$184/10^{6}gal. The total cost of capital, operation, and maintenance for a 3 mgd facility was estimated at \$150/10^6 gal.

Alum was recently evaluated by Shindala and Stewart (13) for phosphorus removal and polishing of waste stabilization pond effluents in Mississippi. Jar tests were used in the evaluation. Using a minimum of 90% removal of phosphorus and 70% removal of COD as criteria, the optimum alum dosage was 85 mg/l at pH 5.5. With higher doses, up to 97% phosphorus removal and 85% COD removal were observed. Ferric chloride and ferric sulfate were also tried but left a residual iron color in the supernatant.

9.3.7 Moving Bed Filter (14)

Alum and polyelectrolyte were used to treat trickling filter effluent in a proprietary moving bed filter at Liberty Corner, New Jersey. The device, a product of Johns-Manville Products Corp., used a bed of sand which is periodically moved upward in an inclined filter vessel. Clean sand is fed in at the bottom of the filter and floc-laden sand is removed mechanically at the top as required by pressure drop, leaving a fresh filtration surface. The wastewater being filtered passes down through the bed and flows through screens at the sides of the filter body.

A major factor controlling phosphate removal is the molar ratio of Al:P. At an alum dosage of 200 mg/l and influent total phosphorus concentrations on the order of 25 to 28 mg/l as P (Al:P ratio = 0.6 to 0.7), the total phosphorus removal efficiency averaged 90%. With lower total phosphorus concentrations (Al:P ratios = 1.2 to 2.6) removal efficiencies averaged 95% and reached as high as 99%. The alum treatment also removed about 70% of suspended solids leaving an average of 15 mg/l in the effluent.

Capital costs for a 1 mgd moving bed filter are estimated at $264,000. Total operating costs for such an installation using 200 mg/l alum are estimated at 12¢/1,000 gal. treated.

9.4 References – Chapter 9

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Chapter 10

STORAGE AND FEEDING OF CHEMICALS

10.1 Aluminum Compounds

The principal aluminum compounds that are commercially available and suitable for phosphorus precipitation are alum and sodium aluminate. Both of these chemicals are available in either liquid or dry forms. Alum is acidic in nature while sodium aluminate is alkaline and therefore one may have an advantage over the other depending on the alkalinity of the wastewater.

10.1.1 Dry Alum

Properties and Availability. The commercial dry alum most often used in wastewater treatment is known as "filter alum" and has the approximate chemical formula $Al_2(SO_4)_3 \cdot 14H_2O$ and a molecular weight of 594. Alum is white to cream in color and a 6% solution has a pH of about 3.4. The commercially available grades of alum and their corresponding bulk densities and angles of repose are:

Grade	Bulk Density (lb/ft^3)	Angle of Repose
Lump	62 to 68	
Ground	60 to 71	43°
Rice	57 to 61	38°
Powdered	38 to 45	65°

Each of these grades has a minimum aluminum content of 17%, expressed as Al_2O_3. Viscosity and solution crystallization temperatures are included in the subsequent section on liquid alum.

Since dry alum is only partially hydrated, it is slightly hygroscopic. However, it is relatively stable when stored under the extremes of temperature and humidity encountered in the United States.

Dry alum is not corrosive unless it absorbs moisture from the air, such as during prolonged exposure to humid atmospheres. Therefore, precautions should be taken to ensure that the storage space is free of moisture.

Alum is shipped in 50 or 100 lb bags, drums, or in bulk by truck or rail. Bag shipments may be ordered on wood pallets if desired. Locations of the major production plants are listed in Table 10-1. At present, the price range for dry alum in bulk quantities is $58 to $62/ton, F.O.B. the point of manufacture. Freight costs to the point of usage must be added to this.

Table 10-1

LOCATION OF ALUM MANUFACTURING PLANTS

Location	Manufacturer	Form of Alum Available
ALABAMA		
Coosa Pines	American Cyanamid	Liquid
Demopolis	American Cyanamid	Liquid
Mobile	American Cyanamid	Liquid and Dry
Naheola	Stauffer	Liquid
ARKANSAS		
Pine Bluff	Allied	Liquid
CALIFORNIA		
Bay Point (San Francisco)	Allied	Liquid and Dry
El Segundo (Los Angeles)	Allied	Liquid
Richmond (San Francisco)	Stauffer	Liquid
Vernon (Los Angeles)	Stauffer	Liquid
COLORADO		
Denver	Allied	Liquid and Dry
FLORIDA		
Fernandina Beach	Tennessee Corp.	Liquid
Jacksonville	Allied	Liquid
Port St. Joe	Allied	Liquid
GEORGIA		
Atlanta (2 plants)	Burris, Allied	Liquid and Dry
Augusta	Tennessee Corp.	Liquid and Dry
Cedar Springs	Tennessee Corp.	Liquid
Macon	Allied	Liquid
Savannah	Allied	Liquid
ILLINOIS		
E. St. Louis	Allied	Liquid and Dry
Joliet	American Cyanamid	Liquid and Dry
LOUISIANA		
Bastrop	Stauffer	Liquid
Baton Rouge	Stauffer	Liquid
Monroe	Allied	Liquid
New Orleans	Allied	Liquid and Dry
Springhill	Stauffer	Liquid
MAINE		
Searsport	Northern	Liquid and Dry
MARYLAND		
Baltimore	Olin	Dry

Table 10-1 (Continued)

Location	Manufacturer	Form of Alum Available
MASSACHUSETTS		
Adams	Holland	Liquid
Salem	Hamblet & Hayes	Liquid and Dry
MICHIGAN		
Detroit	Allied	Liquid
Escanaba	American Cyanamid	Liquid
Kalamazoo (2 plants)	Allied, American Cyanamid	Liquid
MINNESOTA		
Cloquet	American Cyanamid	Liquid
Pine Bend	North Star	Liquid and Dry
MISSISSIPPI		
Monticello	American Cyanamid	Liquid
Vicksburg	Allied	Liquid
NEW JERSEY		
Newark	Essex	Liquid and Dry
Warners	American Cyanamid	Liquid and Dry
NORTH CAROLINA		
Acme	Wright	Liquid
Plymouth	American Cyanamid	Liquid
OHIO		
Chillicothe	Allied	Liquid
Cleveland	Allied	Liquid and Dry
Hamilton	American Cyanamid	Liquid and Dry
Middletown	Allied	Liquid
OREGON		
North Portland	Stauffer	Liquid and Dry
PENNSYLVANIA		
Johnsonburg	Allied	Liquid
Marcus Hook	Allied	Liquid and Dry
Newell	Allied	Liquid
SOUTH CAROLINA		
Catawba	Burris	Liquid
Georgetown	American Cyanamid	Liquid and Dry
TENNESSEE		
Chattanooga	American Cyanamid	Liquid and Dry
Counce	Stauffer	Liquid
Springfield	Burris	Liquid

Table 10-1 (Continued)

Location	Manufacturer	Form of Alum Available
TEXAS		
Houston (2 plants)	Stauffer, Ethyl	Liquid and Dry
VIRGINIA		
Covington	Allied	Liquid
Hopewell	Allied	Liquid
Norfolk	Howerton Gowen	Liquid
WASHINGTON		
Kennewick	Allied	Liquid
Tacoma (2 plants)	Stauffer, Allied	Liquid
Vancouver	Allied	Liquid and Dry
WISCONSIN		
Menasha	Allied	Liquid
Wisconsin Rapids	Allied	Liquid

Manufacturers and Addresses

Allied Chemical Corporation
Industrial Chemicals Division
P. O. Box 1139R
Morristown, New Jersey 07960

American Cyanamid Company
Ind. Chem. Div.
P. O. Box 66189
Chicago, Illinois 60666

Burris Chemical Company
Charleston, South Carolina

Essex Chemical Corporation
1402 Broad Street
Clifton, New Jersey 07015

Ethyl Corporation
Houston, Texas

Hamblet & Hayes Company
Colonial Road
Salem, Massachusetts 10970

Holland Chemical Company
Adams, Massachusetts

Howerton Gowen Company, Inc.
Norfolk, Virginia

Northern Chemical Industries, Inc.
Searsport, Maine

North Star Chemicals Inc.
P. O. Box 28-T
South St. Paul, Minnesota

Olin Corporation,
Chemicals Division
745 Fifth Avenue
New York, New York

Stauffer Chemical Company
299 Park Avenue
New York, New York 10017

Tennessee Corporation
Cities Service Company
Industrial Chemicals Division
P. O. Box 50360
Atlanta, Georgia

Wright Chemical Co.
Acme, North Carolina

General Design Considerations. Ground and rice alum are the grades most commonly used by utilities because of their superior flow characteristics. These grades have less tendency to lump or arch in storage and therefore provide more consistent feeding qualities. Hopper agitation is seldom required with these grades, and in fact may be detrimental to feeding because of the possibility of packing the bin.

Alum dust is present in the ground grade and will cause minor irritation of the eyes and nose on breathing. A respirator may be worn for protection against alum dust. Gloves may be worn to protect the hands. Because of minor irritation in handling and the possibility of alum dust causing rusting of adjacent machinery, dust removal equipment is desirable. Alum dust should be thoroughly flushed from the eyes immediately and washed from the skin with water.

Storage. A typical storage and feeding system for dry alum is shown in Figure 10-1. Bulk alum can be stored in mild steel or concrete bins with dust collector vents located in, above, or adjacent to the equipment room. Dry alum in bulk can be transferred with screw conveyors, pneumatic conveyors, or bucket elevators made of mild steel. Pneumatic conveyor elbows should have a reinforced backing as the alum can contain abrasive impurities.

Bags and drums of alum should be stored in a dry location to avoid caking. Bag or drum loaded hoppers should have a nominal storage capacity for eight hours at the nominal maximum feed rate so that personnel are not required to charge the hopper more than once per shift. Converging hopper sections should have a minimum slope of 60° to prevent arching.

Bulk storage hoppers should terminate at a bin gate so that the feeding equipment may be isolated for servicing. The bin gate should be followed by a flexible connection, and transition hopper chute or hopper which acts as a conditioning chamber over the feeder.

Feeding Equipment. The feed system includes all of the components required for the proper preparation of the chemical solution. Capacities and assemblies should be selected to fulfill individual system requirements. Three basic types of chemical feed equipment are used: volumetric, belt gravimetric, and loss-in-weight gravimetric. Volumetric feeders are usually used where initial low cost and usually lower capacities are the basis of selection. Volumetric feeder mechanisms are usually exposed to the corrosive dissolving chamber vapors which can cause corrosion of discharge areas. Manufacturers usually control this problem by use of an electric heater to keep the feeder housing dry or by using plastic components in the exposed areas.

Volumetric dry feeders presently in general use are of the screw type. Two designs of screw feed mechanism are available. Both allow even withdrawal across the bottom of the feeder hopper to prevent hopper dead zones. One screw design is the variable pitch type with the pitch expanding evenly to the discharge point. The second screw design is the constant pitch-reciprocating type. This type has each half of the screw turned in

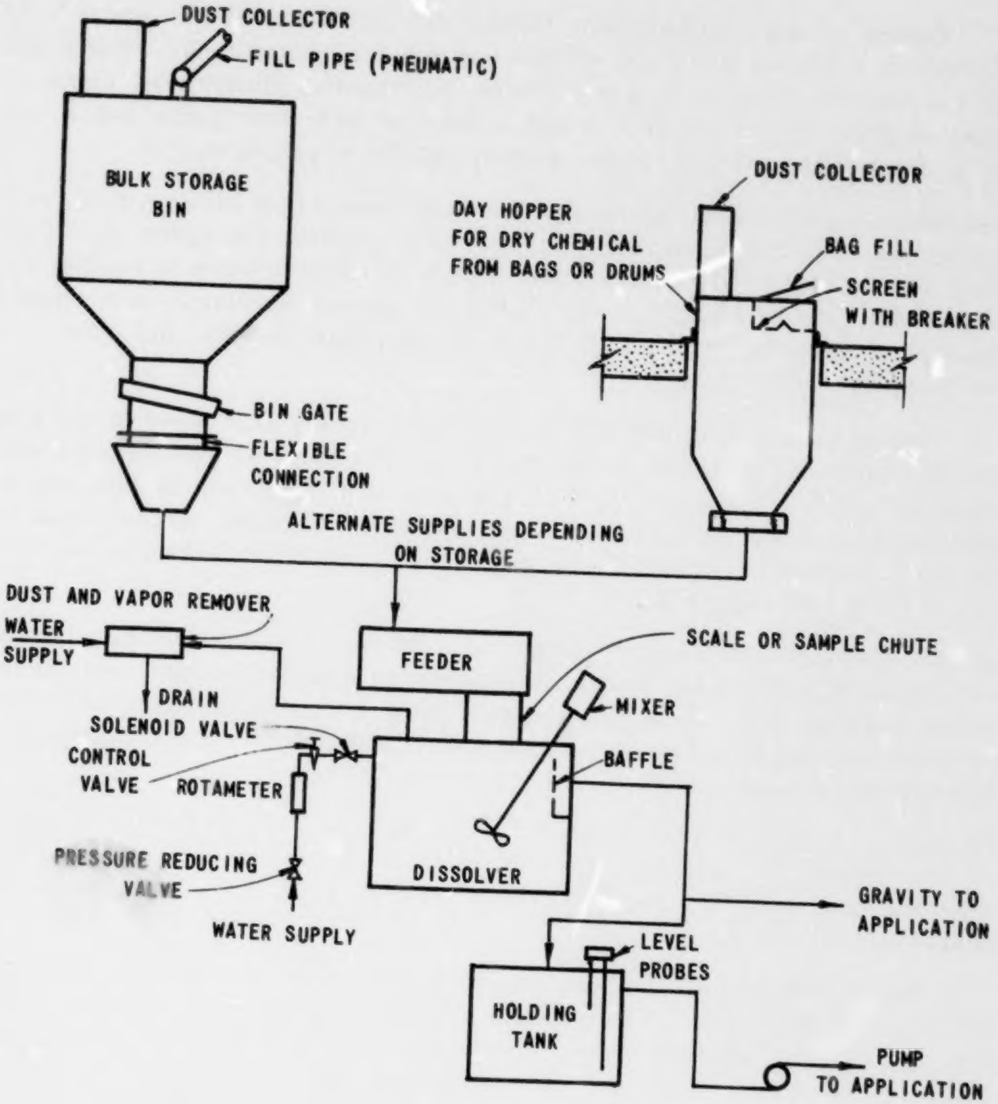

DUST COLLECTOR
FILL PIPE (PNEUMATIC)
BULK STORAGE BIN
BIN GATE
FLEXIBLE CONNECTION
DUST COLLECTOR
DAY HOPPER FOR DRY CHEMICAL FROM BAGS OR DRUMS
BAG FILL
SCREEN WITH BREAKER
ALTERNATE SUPPLIES DEPENDING ON STORAGE
DUST AND VAPOR REMOVER
WATER SUPPLY
DRAIN
FEEDER
SCALE OR SAMPLE CHUTE
MIXER
SOLENOID VALVE
CONTROL VALVE
ROTAMETER
BAFFLE
PRESSURE REDUCING VALVE
DISSOLVER
WATER SUPPLY
GRAVITY TO APPLICATION
LEVEL PROBES
HOLDING TANK
PUMP
TO APPLICATION

FIGURE 10-1 TYPICAL DRY FEED SYSTEM

opposite directions so that the turning and reciprocating motion alternately fills one half of the screw while the other half of the screw is discharging. The variable pitch screw has one point of discharge, while the constant pitch-reciprocating screw has two points of discharge, one at each end of the screw. The accuracy of volumetric feeders is influenced by the character of the material being fed and ranges between ±1% for free-flowing materials and ±7% for cohesive materials. This accuracy is volumetric and should not be related to accuracy by weight (gravimetric).

Where the greatest accuracy and the most economical use of chemical is desired, the loss-in-weight type feeder should be selected. This feeder is limited to the low and intermediate feed rates up to a maximum rate of approximately 4,000 lb/hour. The loss-in-weight type feeder consists of a material hopper and feeding mechanism mounted on enclosed scales. The feed rate controller retracts the scale poise weight to deliver the dry chemical at the desired rate. The feeding mechanism must feed at this rate to maintain the balance of the scale. Any unbalance of the scale beam causes a corrective change in the output of the feeding mechanism. Continuous comparison of actual hopper weight with set hopper weight prevents cumulative errors. Accuracy of the loss-in-weight feeder is ±1% by weight of the set rate.

Belt type gravimetric feeders span the capacity ranges of volumetric and loss-in-weight feeders and can usually be sized for all applications encountered in wastewater treatment applications. Initial expense is greater than for the volumetric feeder and slightly less than for the loss-in-weight feeder. Belt type gravimetric feeders consist of a basic belt feeder incorporating a weighing and control system. Feed rates can be varied by changing either the weight per foot of belt, or the belt speed, or both. Controllers in general use are mechanical, pneumatic, electric and mechanical-vibrating. Accuracy specified for belt type gravimetric feeders should be within ±1% of set rate. Materials of construction of feed equipment normally include mild steel hoppers, stainless steel mechanism components, and rubber surfaced feed belts.

Because alum solution is corrosive, dissolving or solution chambers should be constructed of type 316 stainless steel, fiberglass reinforced plastic (FRP), or plastics. Dissolvers should be sized for preparation of the desired solution strength. The solution strength usually recommended is 0.5 lb of alum to 1 gal. of water, or a 6% solution. The dissolving chamber is designed for a minimum detention time of 5 minutes at the maximum feed rate. Because excessive dilution may be detrimental to coagulation, eductors, or float valves that would ordinarily be used ahead of centrifugal pumps, are not recommended. Dissolvers should be equipped with water meters and mechanical mixers so that the water to alum ratio may be properly established and controlled.

Piping and Accessories. FRP, plastics (polyvinyl chloride, polyethylene, polypropylene, and other similar materials), and rubber are in general use and are recommended for alum solution. Care must be taken to provide adequate support for these piping systems, with close attention given to spans between supports so that objectionable deflection will not be experienced. The alum solution should be injected into a zone of rapid mixing or turbulent flow.

Solution flow by gravity to the point of discharge is desirable. When gravity flow is not possible, transfer components should be selected that require little or no dilution. When metering pumps or proportioning weir tanks are used, return of excess flow to a holding tank should be considered. Metering pumps are discussed further in the section on liquid alum.

Valves used in solution lines should be plastic, type 316 stainless steel or rubber-lined iron or steel.

Pacing and Control. Standard instrument control and pacing signals are generally acceptable for common feeder system operation. Volumetric and gravimetric feeders are usually adaptable to operation from any standard instrument signals.

When solution must be pumped, consideration should be given to use of holding tanks between the dry feed system and feed pumps, and the solution water supply should be controlled to prevent excessive dilution. The dry feeders may be started and stopped from tank level probes. Variable control metering pumps can then transfer the alum stock solution to the point of application without further dilution.

Means should be provided for calibration of the chemical feeders. Volumetric feeders may be mounted on platform scales. Belt feeders should include a sample chute and box to catch samples for checking actual delivery with set delivery.

Gravimetric feeders are usually furnished with totalizers only. Remote instrumentation is frequently used with gravimetric equipment, but seldom used with volumetric equipment.

10.1.2 Liquid Alum

Properties and Availability. Liquid alum is shipped in rubber lined or stainless steel, insulated tank cars or trucks. Alum shipped during the winter is heated prior to shipment so that crystallization will not occur during transit. Liquid Alum is shipped at a solution strength of about 8.3% as Al_2O_3 or about 49% as $Al_2(SO_4)_3 \cdot 14H_2O$. This solution has a density of 11.1 lb/gal. at 60°F and contains about 5.4 lb dry alum (17% Al_2O_3) per gal. of liquid. This solution will begin to crystallize at 18°F.

Crystallization temperatures of other solution strengths are as follows (1):

% Al_2O_3	Temperature of Crystallization, °F.
5.19	26
6.42	21
6.67	19
6.91	17
7.16	15
7.40	13
7.66	12
7.92	14
8.19	17
8.46	20
8.74	28

The viscosity of various alum solutions is given in Figure 10-2.

Tank truck lots of 3,000 to 5,000 gal. are available. Tank car lots are available in quantities of 7,000 to 18,000 gal. Production locations of liquid alum are listed in Table 10-1. The current price range of liquid alum on an equivalent dry alum (17% Al_2O_3) basis is about $43 to $48/ton, F.O.B. the point of manufacture. Liquid alum will generally be more economical than dry alum if the point of use is within a 50 to 100 mile radius of the manufacturing plant.

General Design Considerations. Bulk unloading facilities usually must be provided at the treatment plant. Rail cars are constructed for top unloading and therefore require an air supply system and flexible connectors to pneumatically displace the alum from the car. U.S. Department of Transportation regulations concerning chemical tank car unloading should be observed. Tank truck unloading is usually accomplished by gravity or by a truck mounted pump.

Established practice in the treatment field has been to dilute liquid alum prior to application. However, recent studies have shown that feeding undiluted liquid alum results in better coagulation and settling. This is reportedly due to prevention of hydrolysis of the alum.

No particular industrial hazards are encountered in handling liquid alum. However, a face shield and gloves should be worn around leaking equipment. The eyes or skin should be flushed and washed upon contact with liquid alum. Liquid alum becomes very slick upon evaporation and therefore spillage should be avoided.

Storage. Liquid alum is stored without dilution at the shipping concentration. Storage tanks may be open if indoors but must be closed and vented if outdoors. Outdoor tanks should also be heated, if necessary, to keep the temperature above 18^{o} F to prevent crystallization. Storage tanks should be constructed of type 316 stainless steel; FRP; steel lined with rubber, polyvinyl chloride, or lead. Liquid alum can be stored indefinitely without deterioration.

Storage tanks should be sized according to maximum feed rate, shipping time required, and quantity of shipment. Tanks should generally be sized for 1½ times the quantity of shipments. A 10 day to 2 week supply should be provided to allow for unforeseen shipping delays.

Feeding Equipment. Various types of gravity or pressure feeding and metering units are available. Figures 10-3 and 10-4 illustrate commonly used feed systems. The rotodip-type feeder or rotameter is often used for gravity feed and the metering pump for pressure feed systems.

The pressure or head available at the point of application frequently determines the feeding system to be used. The rotodip feeder can be supplied from overhead storage by gravity with the use of an internal level control valve, as shown by Figure 10-3. It may also be supplied by a centrifugal pump. The latter arrangement requires an excess

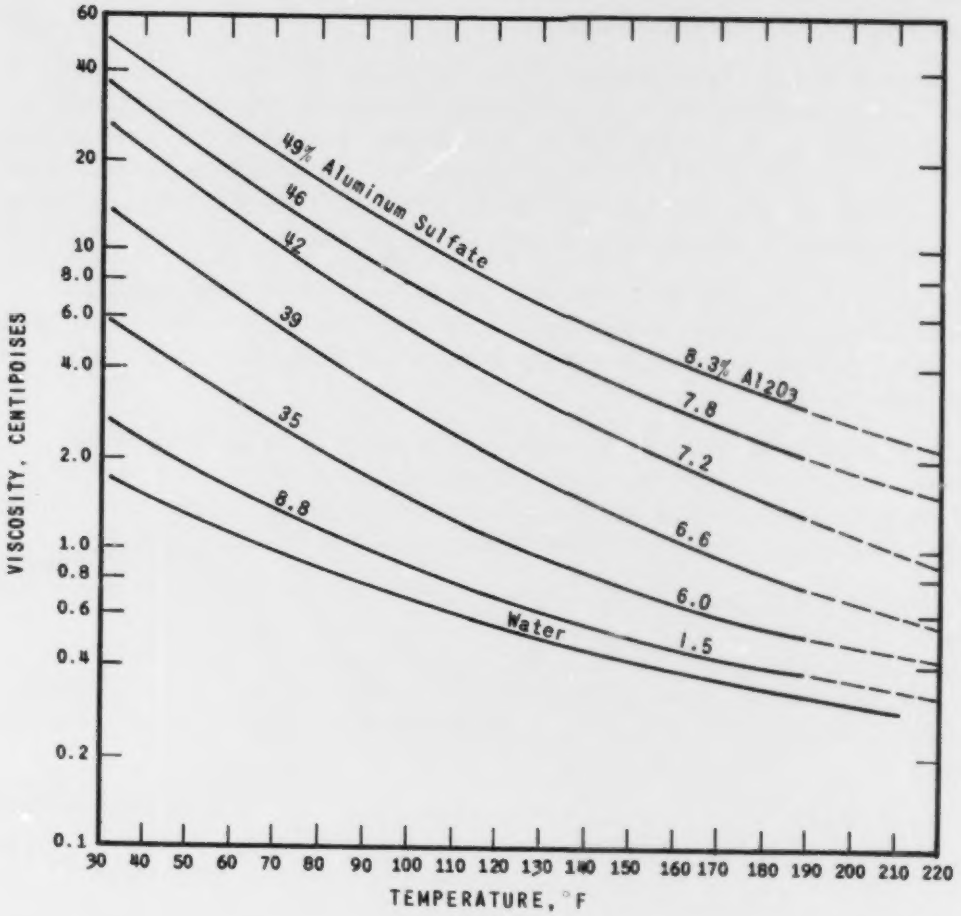


FIGURE 10-2 VISCOSITY OF ALUM SOLUTIONS
(Courtesy of Allied Chemical Co.)

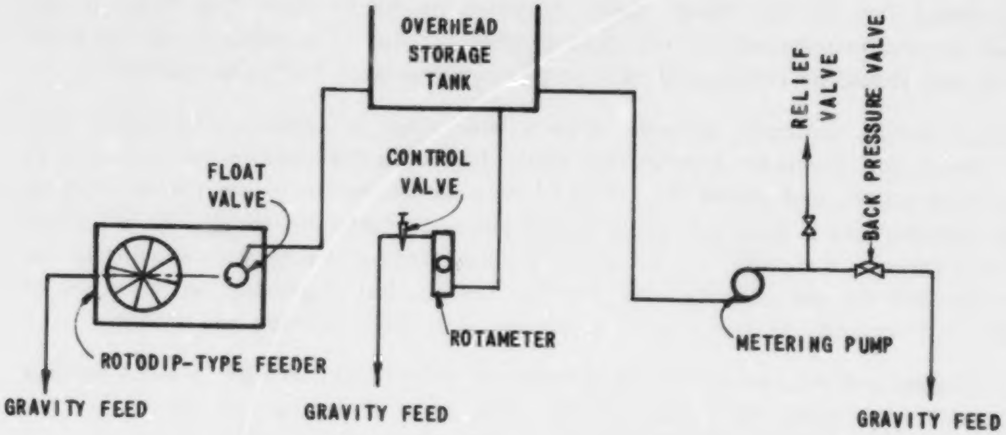

OVERHEAD STORAGE TANK
RELIEF VALVE
BACK PRESSURE VALVE
FLOAT VALVE
CONTROL VALVE
ROTAMETER
METERING PUMP
ROTODIP-TYPE FEEDER
GRAVITY FEED
GRAVITY FEED
GRAVITY FEED

FIGURE 10-3 ALTERNATIVE LIQUID FEED SYSTEMS FOR OVERHEAD STORAGE

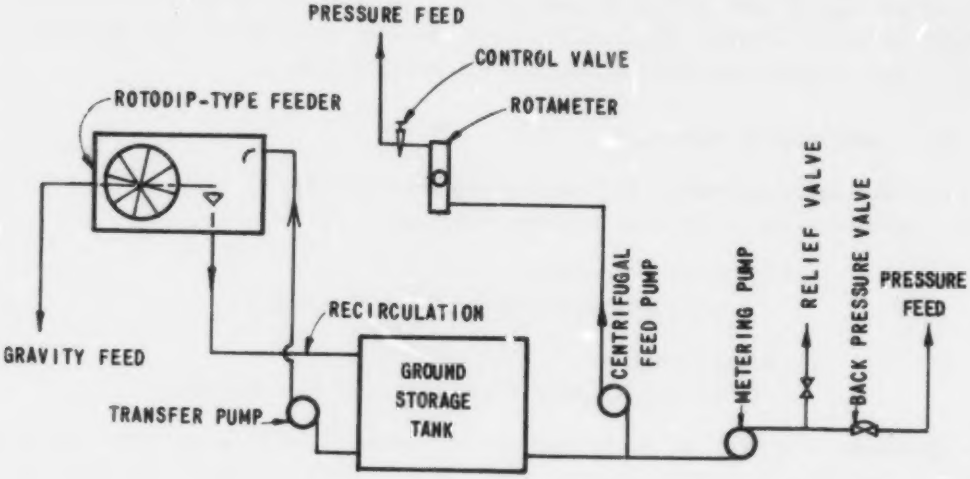

PRESSURE FEED
CONTROL VALVE
ROTAMETER
ROTODIP-TYPE FEEDER
RELIEF VALVE
BACK PRESSURE VALVE
CENTRIFUGAL FEED PUMP
METERING PUMP
PRESSURE FEED
RECIRCULATION
GRAVITY FEED
GROUND STORAGE TANK
TRANSFER PUMP

FIGURE 10-4 ALTERNATIVE LIQUID FEED SYSTEMS FOR GROUND STORAGE

flow return line to the storage tank, as shown by Figure 10-4. Centrifugal pumps should be direct-connected but not close-coupled because of possible leakage into the motor, and should be constructed of type 316 stainless steel, FRP, and plastics.

Metering pumps, currently available, allow a wide range of capacity to compare with the rotodip and rotameter systems. Hydraulic diaphragm type pumps are preferable to other type pumps and should be protected with an internal or external relief valve. A back pressure valve is usually required in the pump discharge to provide efficient check valve action. Materials of construction for feeding equipment should be as recommended by the manufacturer for the service, but depending on the type of system, will generally include type 316 stainless steel, FRP, plastics, and rubber.

Piping and Accessories. Piping systems for alum should be FRP, plastics (subject to temperature limits), type 316 stainless steel, or lead. Piping and valves used for alum solutions are also discussed in the preceding section on dry alum.

Pacing and Control. The feeding systems described above are volumetric, and the feeders generally available can be adapted to receive standard instrument pacing signals. The signals can be used to vary motor speed, variable speed transmission setting, stroke speed and stroke length where applicable. A totalizer is usually furnished with a rotodip type feeder, and remote instruments are available. Instrumentation is rarely used with rotameters and metering pumps.

10.1.3 Dry Sodium Aluminate

Properties and Availability. Dry sodium aluminate, $Na_2Al_2O_4$, is available from two major manufacturers in the United States. They are:

Nalco Chemical Company
Chicago, Illinois (Plant located at Clearing, Ill.)

Reynolds Chemicals
Richmond, Virginia (Plant located at Bauxite, Ark.)

Sodium aluminate is shipped in 50 lb bags and has a bulk density ranging from 40 to 50 lb/ft^3. The Al_2O_3 content ranges from 41 to 46%. Sodium aluminate is noncorrosive and the pH of a 1% solution is about 11.9. Manufacturers should be consulted for more precise specifications of their product.

The current price of dry sodium aluminate in 40,000 lb shipments ranges from $0.11 to $0.14/lb, F.O.B. the point of manufacture. Prices increase substantially for smaller shipments.

General Design Considerations. Requirements for dry sodium aluminate feed systems are generally similar to those for dry aluminum sulfate. Dry sodium aluminate is not available in bulk quantities. Therefore, the small, day type hoppers with manual filling arrangements as shown by Figure 10-1 are used. Precautionary measures for handling sodium aluminate are similar to those for strong alkalies, such as caustic soda.

Contact with skin, eyes, and clothing should be avoided. Aluminate dust or solution spray should not be breathed.

Storage. Dry sodium aluminate is stored as received, in bags, and at optimum conditions of 60 to 90° F, the recommended storage limit is 6 months. Hopper material of mild steel is completely adequate. This chemical may or may not be free flowing, depending on the manufacturer and grade used. Therefore, hopper agitation may be required. Sodium aluminate deteriorates on exposure to the atmosphere and care should be taken to avoid tearing of bags.

Feeding Equipment. Materials of construction for dissolving chambers may be mild steel or stainless steel and selection may be influenced by conformity with adjacent equipment. Equipment similar to that shown by Figure 10-1 is applicable. Standard practice for the free-flowing grade of sodium aluminate calls for dissolvers sized for 0.5 lb/gal. or 6% solution strength with a dissolver detention time of 5 minutes at the maximum feed rate.

After dissolving sodium aluminate in the preparation of batch solutions, agitation should be minimized or eliminated to prevent deterioration of the solution. Air agitation is not recommended, and solution tanks should be covered to prevent carbonation of the solution.

Piping and Accessories. Materials for piping and transporting sodium aluminate solution may be mild steel, iron, type 304 stainless steel, concrete, or plastics. The use of copper, copper alloys, and rubber should be avoided.

Pacing and Control. Pacing and control fundamentals are similar to those described for dry alum.

The amount of dilution does not appear to be a consideration in the use of sodium aluminate. Therefore, the use of float valves to satisfy centrifugal pump suction, and the use of eductors are permissible.

10.1.4 Liquid Sodium Aluminate

Properties and Availability. Liquid sodium aluminate is available from the following major manufacturers in the United States:

Nalco Chemical Company
Chicago, Illinois

Vinings Chemical Company
Vinings, Georgia

Conservation Chemical Company
Kansas City, Missouri

There is considerable variety in the composition of sodium aluminate from the manufacturers listed. The Al_2O_3 content varies from 4.9 to 26.7%. The lower solution strengths are usually more expensive because of the cost of freighting the solution water. Because of the variety of solution strengths available, the manufacturers should be contacted for more specific information on density, viscosity, and cost.

Liquid sodium aluminate is available in 30 gal. drums (380 lb), tank truck, and tank car quantities. The current price of sodium aluminate in 40,000 lb quantities is about $0.08/lb ($Al_2O_3$ content, 26.4%) F.O.B. the point of manufacture.

General Design Considerations. Material selection and dilution restrictions are not as limited as for liquid alum, because of the noncorrosive nature of sodium aluminate. Sodium aluminate is a strong alkali and the same precautions should be exercised in handling it as in handling caustic soda.

Storage. Liquid sodium aluminate is usually stored at the shipping concentration, either in the shipping drums or in mild steel tanks. Storage tanks may be located indoors or outdoors, however, outdoor tanks should be provided with facilities for indirect heating. The maximum recommended length of storage is two months. Bulk shipments can be unloaded by gravity, pumping, or air pressure. However, if air is used, it should first be passed through lime-caustic soda breathers to remove the CO_2. Steam injection facilities are required at the unloading site.

Feeding Equipment. Feeding equipment and systems as described for liquid alum generally apply to sodium aluminate except with changes of requirements regarding dilution and materials of construction as described above.

Liquid sodium aluminate may be fed at shipping strength or diluted to a stable 5 to 10% solution. Stable solutions are prepared by direct addition of low hardness water and mild agitation. Air agitation is not recommended.

Piping and Accessories. Material requirements are the same as previously indicated for solutions of dry aluminate.

Pacing and Control. System pacing and control requirements are the same as described for liquid alum.

10.1.5 Estimated Initial Costs of Adding Liquid Alum Feed Facilities to Existing Wastewater Treatment Plants

Cost estimates of chemical storage and feeding equipment for 1, 10, and 100 mgd plants have been prepared based on the use of liquid alum (8.3% Al_2O_3) for phosphorus precipitation. A mole ratio of Al:P of 2.0:1 was assumed to treat an influent phosphorus concentration of 10 mg/l as P. This corresponds to an alum dose of 191 mg/l (17.4 mg/l as Al^{3+}).

Chemical feed equipment was sized for a peak feed rate of twice that calculated from the mole ratio. Storage was provided for at least 15 days at the average feed rate. This storage time is arbitrary and will vary at each installation depending on the distance to and the reliability of the source of chemical supply. Piping and buildings to house the feeding equipment are not included in the estimates. At many locations the chemical feeding equipment can be installed in existing buildings. In all estimates, the costs include an allowance for the contractor's installation, overhead, and profit, and an allowance of 20% of the construction cost for engineering and contingencies. The estimated costs for each size plant are:

Plant Size (mgd)	Estimated Cost ($)
1	15,000
10	47,500
100	588,000

For the 1.0 mgd plant, feeding equipment for liquid alum includes two 25 gph hydraulic diaphragm pumps (one operating and one standby) with the necessary accessories and equipment to pace the feed rate with plant flow. Two 3,000 gal. FRP tanks with accessories are provided for storage of liquid alum. The total capacity of 6,000 gal. allows some flexibility in ordering and receiving 4,000 gal. tank truck shipments.

For the 10 mgd plant, liquid alum is fed by three 125 gph rotodip-type feeders (two operating and one standby) with the necessary accessories and control panel for proportioning chemical feed to flow. Alum is stored in four 11,500 gal. FRP tanks.

For the 100 mgd plant, liquid alum is fed with seven 415 gph rotodip-type feeders (six operating and one standby) with the necessary control equipment. Storage costs are based on ten 50,000 gal. underground concrete tanks with a rubber lining. The underground storage necessitates transfer pumps and day tanks for the rotodip feeders but eliminates the need for heating the tanks. Three 500 gal. FRP day tanks are included in the estimate; one for each pair of feeders. Four alum transfer pumps (three operating and one standby), with a capacity of 50 gpm at 50 ft of head are provided.

The cost of feeding and storage facilities for the 100 mgd plant is greater than 10 times the cost of facilities for the 10 mgd plant. This is because underground storage facilities were the basis of design for the 100 mgd plant in contrast to FRP tanks located in an existing building for the 10 mgd plant. The cost of a building for the 1.0 and 10.0 mgd plants would raise the cost substantially. FRP tanks could be used for the larger plant, however, structural design may necessitate special construction requirements for underground FRP tanks. FRP tanks located above ground and on concrete foundations would require special insulation and heating.

10.2 Iron Compounds

10.2.1 Liquid Ferric Chloride

Properties and Availability. Liquid ferric chloride is a corrosive, dark brown oily-appearing solution having a density as shipped and stored of 11.2 to 12.4 lb/gal. (35 to 45% $FeCl_3$) (2). The ferric chloride content of these solutions, as $FeCl_3$, is 3.95 to 5.58 lb/gal. Shipping concentrations vary from summer to winter due to the relatively high crystallization temperature of the more concentrated solutions as shown by Figure 10-5. The pH of a 1% solution is 2.0.

The molecular weight of ferric chloride is 162.22. Viscosities of ferric chloride solutions at various temperatures are presented in Figure 10-6.

Liquid ferric chloride is shipped in 3,000 to 4,000 gal. bulk truckload lots, in 4,000 to 10,000 gal. bulk carload lots, and in 5 and 13 gal. carboys. Liquid ferric chloride is produced at the following locations:

Dow Chemical Co.
Midland, Michigan

Pennwalt Corp.
Philadelphia, Pa. (Plant at Wyandotte, Mich.)

The current price of liquid ferric chloride in bulk quantities is about $0.04 to $0.045/lb (as $FeCl_3$), F.O.B. the point of manufacture.

Tank trucks and cars are normally unloaded pneumatically, and operating procedures must be closely followed to avoid spills and accidents. The safety vent cap and assembly (painted red) should be removed prior to opening the unloading connection to depressurize the tank car or truck, prior to unloading.

General Design Considerations. Ferric chloride solutions are corrosive to many common materials and cause stains which are difficult to remove. Areas which are subject to staining should be protected with resistant paint or rubber mats.

Normal precautions should be employed when cleaning ferric chloride handling equipment. Workmen should wear rubber gloves, rubber apron, and goggles or a face shield. If ferric chloride comes in contact with the eyes or skin, flush with copious quantities of running water and call a physician. If ferric chloride is ingested, induce vomiting and call a physician.

Storage. Ferric chloride solution can be stored as shipped. Storage tanks should have a free vent or vacuum relief valve. Tanks may be constructed of FRP, rubber lined steel, or plastic lined steel. Resin impregnated carbon or graphite are also suitable materials for storage containers.

It may be necessary in most instances to house liquid ferric chloride tanks in heated areas or provide tank heaters or insulation to prevent crystallization. Ferric chloride can be stored for long periods of time without deterioration. The total storage capacity should be 1½ times the largest anticipated shipment, and should provide at least a 10 day to 2 week supply of the chemical at the design average dosage.

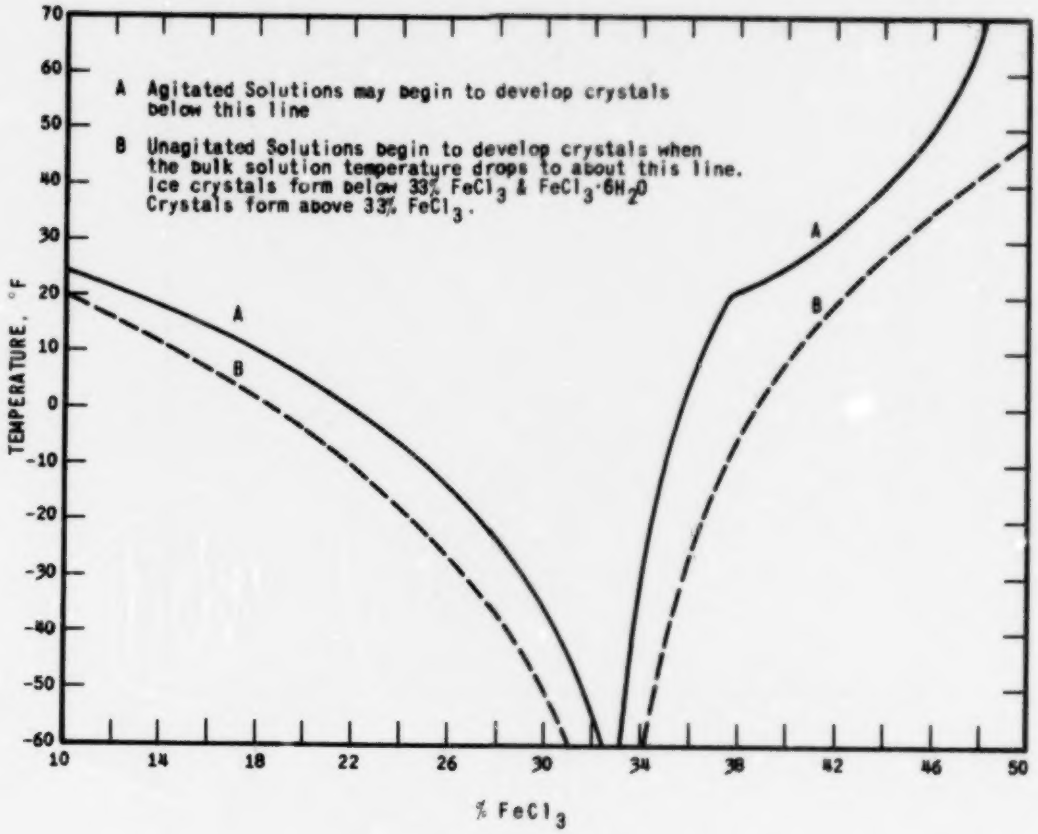


FIGURE 10-5 FREEZING POINT CURVES FOR COMMERCIAL FERRIC CHLORIDE SOLUTIONS
(Courtesy of Dow Chemical Co.)

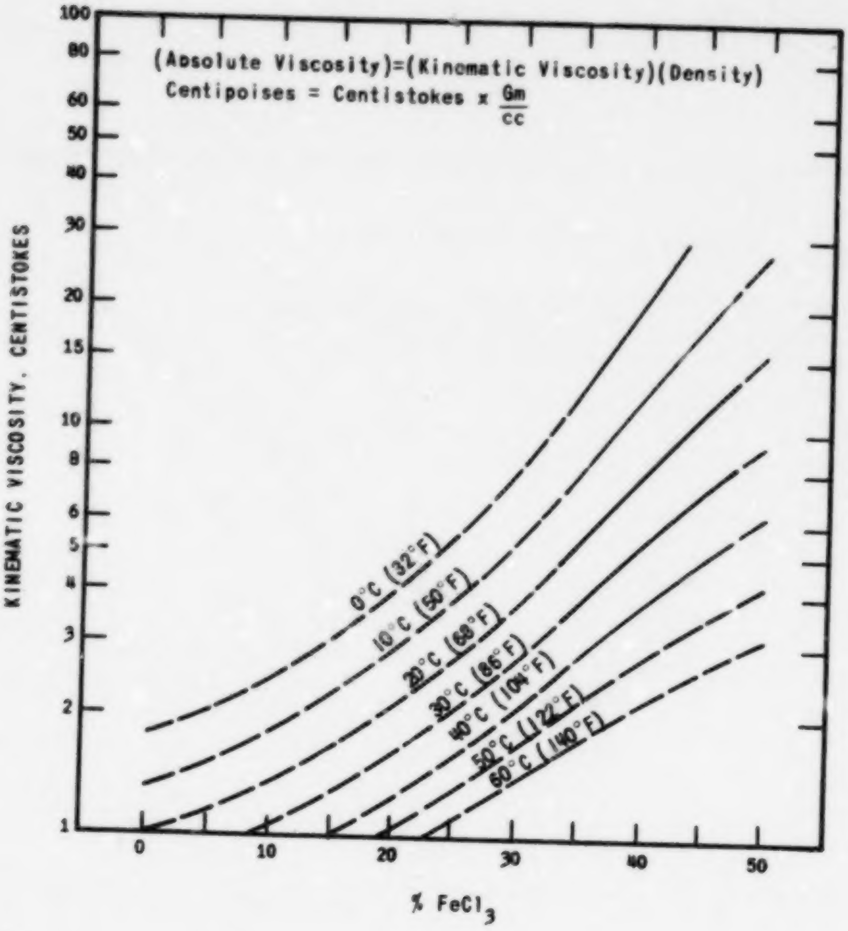


FIGURE 10-6 VISCOSITY VS COMPOSITION OF FERRIC CHLORIDE SOLUTIONS AT VARIOUS TEMPERATURES
(Courtesy of Dow Chemical Co.)

Feeding Equipment. Feeding equipment and systems described for liquid alum generally apply to ferric chloride except for materials of construction, and the use of glass tube rotameters.

It may not be desirable to dilute the ferric chloride solution from its shipping concentration to a weaker feed solution because of possible hydrolysis. Ferric chloride solutions may be transferred from underground storage to day tanks with impervious graphite or rubber lined self-priming centrifugal pumps having teflon rotary and stationary seals. Because of the tendency for liquid ferric chloride to stain or deposit, glass-tube rotameters should not be used for metering this solution. Rotodip feeders and diaphragm metering pumps are often used for ferric chloride, and should be constructed of materials such as rubber lined steel and plastics.

Piping and Accessories. Materials for piping and transporting ferric chloride should be rubber or Saran lined steel, hard rubber, FRP, or plastics. Valving should consist of rubber or resin lined diaphragm valves, Saran lined valves with teflon diaphragms, rubber sleeved pinch type valves, or plastic ball valves. Gasket material for large openings such as manholes in storage tanks should be soft rubber; all other gaskets should be graphite-impregnated blue asbestos, teflon, or vinyl.

Pacing and Control. System pacing and control requirements are similar to those discussed previously for liquid alum.

10.2.2 Ferrous Chloride (Waste Pickle Liquor)

Properties and Availability. Ferrous chloride, $FeCl_2$, as a liquid is available in the form of waste pickle liquor from steel processing. The liquor weighs between 9.9 and 10.4 lb/gal. and contains 20 to 25% $FeCl_2$ or about 10% available Fe^{2+}. A 22% solution of $FeCl_2$ will crystallize at a temperature of -4° F. The molecular weight of $FeCl_2$ is 126.76. Free acid in waste pickle liquor can vary from 1 to 10% and usually averages about 1.5 to 2.0%. Ferrous chloride is slightly less corrosive than ferric chloride.

Waste pickle liquor is available in 4,000 gal. truckload lots and a variety of carload lots. In most instances the availability of waste pickle liquor will depend on the proximity to steel processing plants. Dow Chemical Company produces a waste pickle liquor, having an $FeCl_2$ content of about 22% at a price of \$0.04/lb of $FeCl_2$ in bulk car or truckload quantities, F.O.B. Midland, Michigan.

General Design Considerations. Since ferrous chloride or waste pickle liquor may not be available on a continuous basis, storage and feeding equipment should be suitable for handling ferric chloride. Therefore, the ferric chloride section should be referred to for storage and handling details.

10.2.3 Ferric Sulfate

Properties and Availability. Ferric sulfate is marketed as a dry, partially hydrated product with a maximum of three water molecules. Density, % $Fe_2(SO_4)_3$, and molecular weight of the two most common hydrates are presented below:

Properties of Two Ferric Sulfate Hydrates (3)

	$Fe_2(SO_4)_3 \cdot 3\ H_2O$	$Fe_2(SO_4)_3 \cdot 2\ H_2O$
Density	63 to 72 lb/ft^3	83 to 97 lb/ft^3
% $Fe_2(SO_4)_3$	68%	76%
Molecular weight	453.8	435.8

Ferric sulfate is shipped in car and truck load lots of 50 lb and 100 lb moistureproof paper bags and 400 lb fiber drums. Bulk carload shipments in box and closed hopper cars are available. Major producers and corresponding plant locations are:

Producer	Plant Location
Stauffer Chemical Co.	San Francisco, California Fort Worth, Texas
Tennessee Corp. (Cities Service Co.)	Copper Hill, Tennessee

The current price of ferric sulfate (21.8% Fe) is about $38/ton, F.O.B. Copper Hill, Tennessee.

General precautions should be observed when handling ferric sulfate, such as wearing goggles and dust masks, and areas of the body that come in contact with the dust or vapor should be washed promptly.

General Design Considerations. Aeration of ferric sulfate should be held to a minimum because of the hygroscopic nature of the material, particularly in damp atmospheres. Mixing of ferric sulfate and quicklime in conveying and dust vent systems should be avoided as caking and excessive heating can result. The presence of ferric sulfate and lime in combination has been known to destroy cloth bags in pneumatic unloading devices (4). Because ferric sulfate in the presence of moisture will stain, precautions similar to those discussed for ferric chloride should be observed.

Storage. Ferric sulfate is usually stored in the dry state either in the shipping bags or in bulk in concrete or steel bins. Bulk storage bins should be as tight as possible to avoid moisture absorption, but dust collector vents are permissible and desirable.

Bins may be located inside or outside and the material transferred by bucket elevator, screw or air conveyors. Ferric sulfate stored in bins usually absorbs some moisture and forms a thin protective crust which retards further absorption until the crust is broken.

Feeding Equipment. Feed solutions are usually made up at a water to chemical ratio of 2:1 to 8:1 (on a weight basis) with the usual ratio being 4:1 with a 20 minute detention time. Care must be taken not to dilute ferric sulfate solutions to less than 1% to prevent hydrolysis and deposition of ferric hydroxide. Ferric sulfate is actively corrosive in solution, and dissolving and transporting equipment should be fabricated of type 316 stainless steel, rubber, plastics, ceramics or lead.

Dry feeding requirements are similar to those for dry alum except that belt type feeders are rarely used because of their open type of construction. Closed construction, as found in the volumetric and loss-in-weight type feeders, generally exposes a minimum of operating components to the vapor, and thereby minimizes maintenance. A water jet vapor remover should be provided at the dissolver to protect both the machinery and operator.

Piping and Accessories. Piping systems for ferric sulfate should be FRP, plastics, type 316 stainless steel, rubber or glass.

Pacing and Control. System pacing and control are the same as discussed for dry alum.

10.2.4 Ferrous Sulfate

Properties and Availability. Ferrous sulfate or copperas is a byproduct of pickling steel and is produced as granules, crystals, powder, and lumps. The most common commercial form of ferrous sulfate is $FeSO_4 \cdot 7H_2O$, with a molecular weight of 278, and containing 55 to 58% $FeSO_4$ and 20 to 21% Fe. The product has a bulk density of 62 to 66 lb/ft^3. When dissolved, ferrous sulfate is acidic. The composition of ferrous sulfate may be quite variable and should be established by consulting the nearest manufacturers.

Bulk, drum (400 lb) and bag (50 and 100 lb) shipments are available from producers at the following locations:

American Cyanamid Co.	Savannah, Georgia
Byproducts Processing Co, Inc.	Baltimore, Maryland
Glidden Co.	Baltimore, Maryland
Cosmin Corp.	Baltimore, Maryland
National Lead	St. Louis, Missouri
National Lead	Sayreville, New Jersey

The current price of ferrous sulfate in bulk carload and truckload quantities is about $18/ton (21% Fe).

Ferrous sulfate is also available in a wet state in bulk form from some plants. This form is likely to be difficult to handle and the manufacturer should be consulted for specific information and instructions.

Dry ferrous sulfate cakes at storage temperatures above 68° F, is efflorescent in dry air, and oxidizes and hydrates further in moist air.

General precautions similar to those for ferric sulfate with respect to dust and handling acidic solutions, should be observed when working with ferrous sulfate. Mixing quicklime and ferrous sulfate produces high temperatures and the possibility of fire.

General Design Considerations. The granular form of ferrous sulfate has the best feeding characteristics and gravimetric or volumetric feeding equipment may be used.

The optimum chemical to water ratio for continuous dissolving is 0.5 lb/gal. or 6% with a detention time of 5 minutes in the dissolver. Mechanical agitation should be provided in the dissolver to assure complete solution. Lead, rubber, iron, plastics, and type 304 stainless steel can be used as construction materials for handling solutions of ferrous sulfate.

Storage, feeding and transporting systems probably should be suitable for handling ferric sulfate as an alternative to ferrous sulfate.

10.2.5 Estimated Initial Costs of Adding Liquid Ferric Chloride Feed Systems to Existing Wastewater Treatment Plants

Costs of chemical storage and feeding equipment for 1, 10, and 100 mgd plants have been estimated based on the use of liquid ferric chloride (basis: 35% $FeCl_3$) for phosphorus precipitation. A mole ratio of Fe:P of 2.0:1 was assumed to treat an influent phosphorus concentration of 10 mg/l as P. This corresponds to a ferric chloride dose of 104 mg/l (36 mg/l as Fe^{3+}).

Chemical feed equipment was sized for a peak feed rate of twice that calculated from the mole ratio. Storage was provided for at least 15 days at the average feed rate. This storage time is arbitrary and will vary at each installation depending on the distance to and the reliability of the source of chemical supply. Piping and buildings to house the feeding equipment are not included. At many locations the chemical feeding equipment can be installed in existing buildings. All estimated costs include an allowance for the contractor's installation, overhead, and profit plus an allowance of 20% of the construction cost for engineering and contingencies. The estimated costs for each size plant are:

Plant Size (mgd)	Estimated Cost ($)
1	15,000
10	45,000
100	494,000

For the 1 mgd plant, the feeding equipment for liquid ferric chloride includes two 18 gph, hydraulic diaphragm pumps (one operating and one standby) with the necessary accessories and equipment to pace the feed rate to the plant flow. Liquid ferric chloride is stored in two 3,000 gal. FRP tanks with accessories. The 6,000 gal. total capacity allows use of liquid ferric chloride in 4,000 gal. tank truck quantities.

For the 10 mgd plant, liquid ferric chloride is fed by three 90 gph rotodip-type feeders (two operating and one standby) with the necessary accessories and control panel for proportioning chemical feed to flow.

For the 100 mgd plant, seven 305 gph rotodip-type feeders (six operating and one standby) are provided along with the necessary control equipment. Storage is provided in eight 50,000 gal. rubber-lined concrete tanks located underground. The underground storage necessitates transfer pumps and day tanks for the rotodip feeders, but eliminates the need for heating the tanks. Three 500 gal. FRP day tanks are included along with four ferric chloride transfer pumps sized for 35 gpm each at 50 ft of head.

With regard to the cost of the feed system for the 100 mgd plant being out of proportion to that for a 10 mgd plant, the reader is referred to Section 10.1.5 of this chapter for the discussion of underground concrete storage vessels as opposed to FRP tanks.

10.3 Lime

10.3.1 Quicklime

Properties and Availability. Quicklime, CaO, has a density range of approximately 55 to 70 lb/ft^3, and a molecular weight of 56.08. A slurry for feeding, called milk of lime, can be prepared with up to 45% solids. Lime is only slightly soluble, and both lime dust and slurries are caustic in nature. A saturated solution of lime has a pH of about 12.4.

Lime can be purchased in bulk in both car and truck load lots. It is also shipped in 80 and 100 lb multiwall "moistureproof" paper bags. Lime is produced at the locations indicated by Table 10-2.

Table 10-2

LOCATION OF LIME MANUFACTURING PLANTS (6)

Location	Manufacturer	Form of Lime Available
ALABAMA		
Allgood	Cheney Lime & Cement Co.	High Calcium
Keystone	Southern Cement Co., Div. Martin Marietta Corp.	High Calcium
Landmark	Cheney Lime & Cement Co.	High Calcium
Montevallo	U. S. Gypsum Co.	High Calcium
Roberta	Southern Cement Co. Div. Martin Marietta Corp.	High Calcium
Saginaw	Longview Lime Co., Div. Woodward Co., Div. Mead Corp.	High Calcium
Siluria	Alabaster Lime Co.	High Calcium
ARIZONA		
Douglas	Paul Lime Plant, Inc.	High Calcium
Globe	Hoopes & Co.	High Calcium
Nelson	U. S. Lime Div., The Flintkote Co.	High Calcium

Table 10-2 (Continued)

Location	Manufacturer	Form of Lime Available
ARKANSAS		
Batesville	Batesville White Lime Co., Div. Rangaire Corp.	High Calcium
CALIFORNIA		
City of Industry	U. S. Lime Div., The Flintkote Co.	High Calcium
Diamond Springs	Diamond Springs Lime Co.	High Calcium
Lucerne Valley	Pfizer, Inc., Minerals, Pigments and Metals Div.	High Calcium
Richmond	U. S. Lime Div., The Flintkote Co.	High Calcium
Salinas	Kaiser Aluminum & Chemical Corp. (currently captive lime)	Dolomitic
Sonora	U. S. Lime Div., The Flintkote Co.	Dolomitic
Westend	Stauffer Chemical Co.	High Calcium
COLORADO		
Ft. Morgan	Great Western Sugar Co.	High Calcium
CONNECTICUT		
Canaan	Pfizer, Inc., Minerals, Pigments and Metals Div.	Dolomitic
FLORIDA		
Brooksville	Chemical Lime Co.	High Calcium
Sumterville	Dixie Lime and Stone Co.	High Calcium
ILLINOIS		
Marblehead	Marblehead Lime Co.	High Calcium
McCook	Standard Lime & Refractories Div., Martin Marietta Corp.	Dolomitic
Quincy	Marblehead Lime Co.	High Calcium
So. Chicago	Marblehead Lime Co.	High Calcium
Thornton	Marblehead Lime Co.	Dolomitic
INDIANA		
Buffington	Marblehead Lime Co.	High Calcium
IOWA		
Davenport	Linwood Stone Products Co., Inc.	High Calcium
KENTUCKY		
Carntown	Black River Mining Co.	High Calcium
LOUISIANA		
Morgan City	Pelican State Lime Corp.	High Calcium
New Orleans	U. S. Gypsum Co.	High Calcium

Table 10-2 (Continued)

Location	Manufacturer	Form of Lime Available
MARYLAND		
LeGore	LeGore Lime Co.	High Calcium
Woodsboro	S. W. Barrick & Sons, Inc.	High Calcium
MASSACHUSETTS		
Adams	Pfizer, Inc., Minerals, Pigments and Metals Div.	High Calcium
Lee	Lee Lime Corp.	Dolomitic
MICHIGAN		
Detroit	Detroit Lime Co.	High Calcium
Ludington	Dow Chemical Co. (currently captive lime)	High Calcium
Menominee	Limestone Products Co., Div. C. Reiss Coal Co.	High Calcium
River Rouge	Marblehead Lime Co.	High Calcium
MINNESOTA		
Duluth	Cutler Magner Co.	High Calcium
MISSOURI		
Bonne Terre	Valley Dolomite Co.	Dolomitic
Hannibal	Marblehead Lime Co.	High Calcium
Ste. Genevieve	Mississippi Lime Co.	High Calcium
Springfield	Ash Grove Cement Co.	High Calcium
NEVADA		
Apex	U. S. Lime Div., The Flintkote Co.	High Calcium
Henderson	U. S. Lime Div., The Flintkote Co.	Dolomitic & High Calcium
McGill	Morrison-Weatherly Corp.	High Calcium
Sloan	U. S. Lime Div., The Flintkote Co.	Dolomitic & High Calcium
NEW JERSEY		
Newton	Limestone Products Corp. of America	High Calcium
OHIO		
Ashtabula	Union Carbide Olefins Co.	High Calcium
Carey	National Lime & Stone Co.	Dolomitic
Cleveland	Cuyahoga Lime Co.	High Calcium
Delaware	Marble Cliff Quarries Co.	High Calcium
Geona	U. S. Gypsum Co.	Dolomitic
Gibsonburg (2 plants)	Pfizer, Inc., Minerals, Pigments and Metal Div., National Gypsum Co.	Dolomitic
Huron	Huron Lime Co.	High Calcium

Table 10-2 (Continued)

Location	Manufacturer	Form of Lime Available
OHIO (Continued)		
Marble Cliff	Marble Cliff Quarries Co.	High Calcium
Millersville	J. E. Baker Co.	Dolomitic
Woodville	Ohio Lime Co., Standard Lime & Refractories Div., Martin Marietta Corp.	Dolomitic
OKLAHOMA		
Marble City	St. Clair Lime Co.	High Calcium
Sallisaw	St. Clair Lime Co.	High Calcium
OREGON		
Baker	Chemical Lime Co. of Oregon	High Calcium
Portland	Ash Grove Cement Co.	High Calcium
PENNSYLVANIA		
Annville	Bethlehem Mines Corp.	High Calcium
Bellefonte (2 plants)	National Gypsum Co., Warner Co.	High Calcium
Branchton	Mercer Lime & Stone Co.	High Calcium
Devault	Warner Co.	Dolomitic
Everett	New Enterprise Stone & Lime Co.	High Calcium
Pleasant Gap	Standard Lime & Refractories Div., Martin Marietta Corp.	High Calcium
Plymouth Meeting	G. & W. H. Corson, Inc.	Dolomitic
SOUTH DAKOTA		
Rapid City	Pete Lien & Sons, Inc	High Calcium
TENNESSEE		
Knoxville (2 plants)	Foote Mineral Co., Williams Lime Manufacturing Co.	High Calcium
TEXAS		
Blum	Round Rock Lime Companies	High Calcium
Cleburne	Texas Lime Co., Div. Rangaire Corp.	High Calcium
Clifton	Clifstone Lime Co.	High Calcium
Houston	U. S. Gypsum Co.	High Calcium
McNeil	Austin White Lime Co.	High Calcium
New Braunfels	U. S. Gypsum Co.	High Calcium
Round Rock	Round Rock Lime Companies	High Calcium
San Antonio	McDonough Bros., Inc.	High Calcium
UTAH		
Grantsville	U. S. Lime Div., The Flintkote Co.	Dolomitic & High Calcium
Lehi	Rollins Mining Supplies Co.	High Calcium

Table 10-2 (Continued)

Location	Manufacturer	Form of Lime Available
VERMONT		
Winooski	Vermont Assoc. Lime Industries, Inc.	High Calcium
VIRGINIA		
Clearbrook	W. S. Frey Co., Inc.	High Calcium
Kimballton (2 plants)	Foote Mineral Co., National Gypsum Company	High Calcium
Stephens City	M. J. Grove Lime Co., Div. The Flintkote Co.	High Calcium
Strasburg	Chemstone Corp.	High Calcium
WASHINGTON		
Tacoma	Domtar Chemicals Inc.	High Calcium
WEST VIRGINIA		
Millville	Standard Lime & Refractories Div., Martin Marietta Corp.	Dolomitic
Riverton	Germany Valley Limestone Div., Greer Steel Co.	High Calcium
WISCONSIN		
Eden	Western Lime & Cement Co.	Dolomitic
Green Bay	Western Lime & Cement Co.	High Calcium
Knowles	Western Lime & Cement Co.	Dolomitic
Manitowoc	Rockwell Lime Co.	Dolomitic
Superior	Cutler-LaLiberte-McDougall Corp.	High Calcium

Current prices for bulk pebble quicklime range from $17/ton to $21/ton with the higher prices generally in the far west, and higher than average in the north. Bagging adds approximately $4/ton to the cost.

The CaO content of commercially available quicklime can vary quite widely over an approximate range of 70 to 96%. Content below 88% is generally considered below standard in the municipal use field (5). Purchase contracts are often based on 90% CaO content with provisions for payment of a bonus for each 1% over and a penalty for each 1% under the standard. A CaO content less than 75% probably should be rejected because of excessive grit and difficulties in slaking.

Workmen should wear protective clothing and goggles to protect the skin and eyes, as lime dust and hot slurry can cause severe burns. Areas contacted by lime should be washed immediately. Lime should not be mixed with chemicals which have water of hydration. The lime will be slaked by the water of hydration causing excessive temperature rise and possibly explosive conditions. Conveyors and bins used for more than one chemical should be thoroughly cleaned before switching chemicals.

General Design Considerations. Pebble quicklime, all passing a 3/4 in. screen and not more than 5% passing a No. 100 screen, is normally specified because of easier handling and less dust. Hopper agitation is generally not required with the pebble form. Published slaker capacity ratings require "soft or normally burned" limes which provide fast slaking and temperature rise, but poorer grades of limes may also be satisfactorily slaked by selection of the appropriate slaker retention time and capacity.

Storage. Storage of bagged lime should be in a dry place, and preferably elevated on pallets to avoid absorption of moisture. System capacities often make the use of bagged quicklime impractical.

Bulk lime is stored in air-tight concrete or steel bins having a 60° slope on the bin outlet. Bulk lime can be conveyed by conventional bucket elevators and screw, belt, apron, drag-chain, and bulk conveyors of mild steel construction. Pneumatic conveyors subject the lime to air-slaking and particle sizes may be reduced by attrition. Dust collectors should be provided on manually and pneumatically filled bins.

Feeding Equipment. A typical lime storage and feed system is illustrated in Figure 10-7. Quicklime feeders are usually limited to the belt or loss-in-weight gravimetric types because of the wide variation of the bulk density. Feed equipment should have an adjustable feed range of at least 20:1 to match the operating range of the associated slaker. The feeders should have an over-under feed rate alarm to immediately warn of operation beyond set limits of control. The feeder drive should be instrumented to be interrupted in the event of excessive temperature in the slaker compartment.

Lime slakers for wastewater treatment should be of the continuous type, and the major components should include one or more slaking compartments, a dilution compartment, a grit separation compartment and a continuous grit remover. Commercial designs vary in regard to the combination of water to lime, slaking temperature, and slaking time, in obtaining the "milk of lime" suspensions.

The "paste" type slaker admits water as required to maintain a desired mixing viscosity. This viscosity therefore sets the operating retention time of the slaker. The paste slaker usually operates with a low water to lime ratio (approximately 2:1 by weight), elevated temperature, and five minute slaking time at maximum capacity.

The "detention" type slaker admits water to maintain a desired ratio with the lime, and therefore the lime feed rate sets the retention time of the slaker. The detention slaker operates with a wide range of water to lime ratios (2.5:1 and 6:1), moderate temperature, and a 10 minute slaking time at maximum capacity. A water to lime ratio of from 3.5:1 to 4:1 is most often used. The operating temperature in lime slakers is a function of the water to lime ratio, lime quality, heat transfer, and water temperature. Lime slaking evolves heat in hydrating the CaO to $Ca(OH)_2$, and therefore, vapor removers are required for feeder protection.

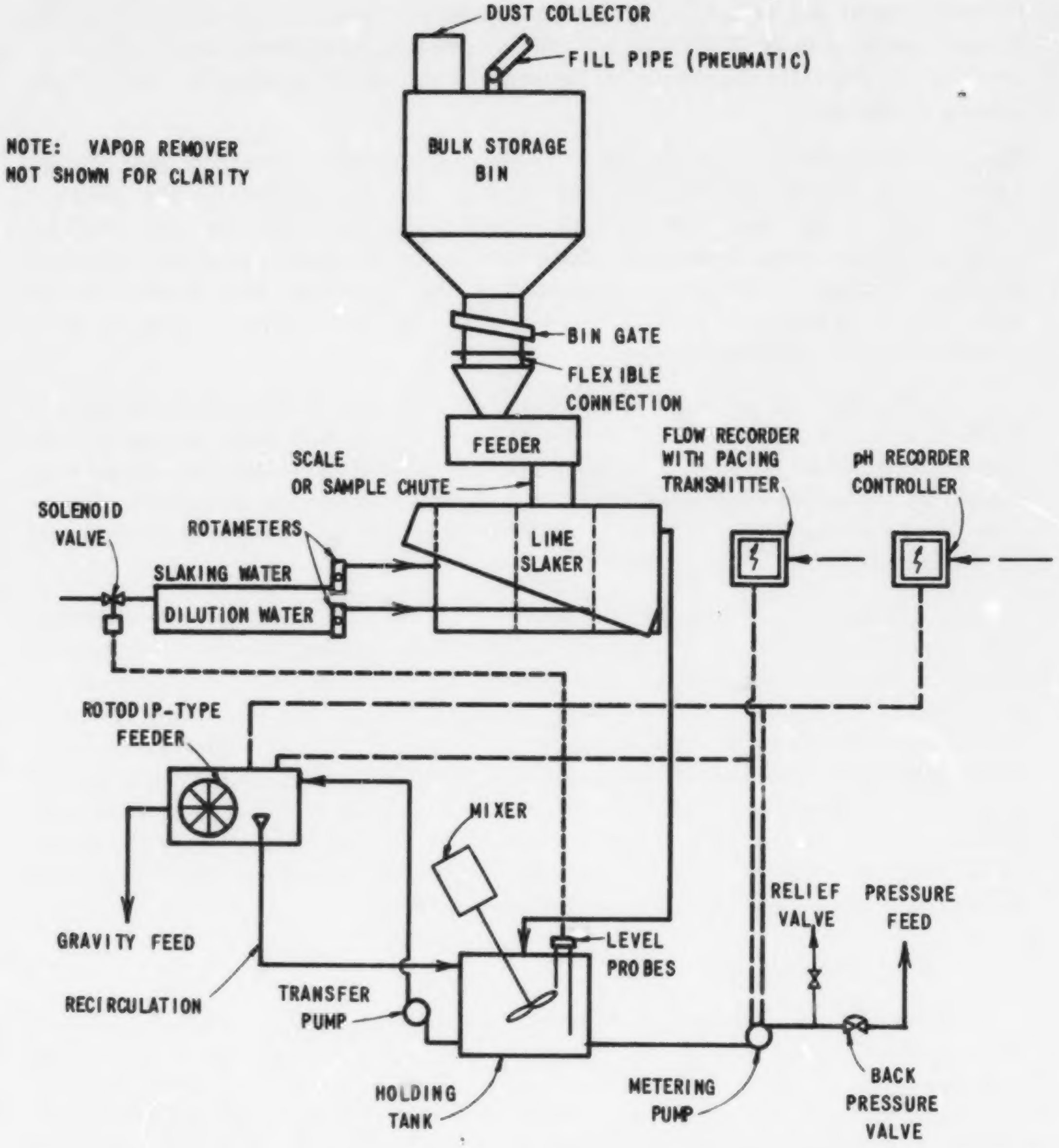

DUST COLLECTOR
FILL PIPE (PNEUMATIC)
NOTE: VAPOR REMOVER
NOT SHOWN FOR CLARITY
BULK STORAGE BIN
BIN GATE
FLEXIBLE CONNECTION
FEEDER
SCALE OR SAMPLE CHUTE
FLOW RECORDER WITH PACING TRANSMITTER
pH RECORDER CONTROLLER
SOLENOID VALVE
ROTAMETERS
SLAKING WATER
DILUTION WATER
LIME SLAKER
ROTODIP-TYPE FEEDER
MIXER
GRAVITY FEED
RECIRCULATION
TRANSFER PUMP
LEVEL PROBES
RELIEF VALVE
PRESSURE FEED
HOLDING TANK
METERING PUMP
BACK PRESSURE VALVE

FIGURE 10-7 TYPICAL LIME FEED SYSTEM

Piping and Accessories. Lime slurry should be transported by gravity in open channels wherever possible. Piping channels, and accessories may be rubber, iron, steel, concrete, and plastics. Glass tubing, such as that in rotameters, will cloud rapidly and therefore should not be used. Any abrupt directional changes in piping should include plugged tees or crosses to allow rodding-out of deposits. Long sweep elbows should be provided to allow the piping to be cleaned by the use of a cleaning "pig". Daily cleaning is desirable.

Milk of lime transfer pumps should be of the open impeller centrifugal type. Pumps having an iron body and impeller with bronze trim are suitable for this purpose. Rubber lined pumps with rubber covered impellers are also frequently used. Make-up tanks are usually provided ahead of centrifugal pumps to ensure a flooded suction at all times. "Plating-out" of lime is minimized by the use of soft water in the make-up tank. Turbine pumps and eductors should be avoided in transferring milk of lime because of scaling problems.

Pacing and Control. Lime slaker water proportioning is integrally controlled or paced from the feeder. Therefore, the feeder-slaker system will follow pacing controls applied to the feeder only. As discussed previously, gravimetric feeders are adaptable to receive most standard instrumentation pacing signals. Systems can be instrumented to allow remote pacing with telemetering of temperature and feed rate to a central panel for control purposes.

The lime feeding system may be controlled by an instrumentation system integrating both plant flow and pH of the wastewater after lime addition. However, it should be recognized that pH probes require daily maintenance in this application to monitor the pH accurately. Deposits tend to build up on the probe and necessitate frequent maintenance. The low pH lime treatment systems (pH 9.5 to 10.0) can be more readily adapted to this method of control than high lime treatment systems (pH 11.0 or greater) because less maintenance of the pH equipment is required. In a closed loop pH-flow control system, milk of lime is prepared on a batch basis and transferred to a holding tank with variable output feeders set by the flow and pH meters to proportion the feed rate. Figure 10-7 illustrates such a control system.

10.3.2 Hydrated Lime

Properties and Availability. Hydrated lime, $Ca(OH)_2$, has a working density of 25 to 35 lb/ft^3, tends to flood the feeder, and will arch in storage bins if packed. The molecular weight is 74.08. The dust and slurry of hydrated lime are caustic in nature. The cost of bulk hydrated lime varies from $18 to $22/ton. Bagged lime is available but increases the cost about $4/ton. The availability of hydrated lime may be determined by contacting manufacturers listed in Table 10-2. The pH of a saturated, hydrated lime solution is the same as that given for quicklime.

General Design Considerations. Hydrated lime is slaked lime and needs only enough water added to form milk of lime. Wetting or dissolving chambers are usually designed to provide 5 minutes detention with a ratio of 0.5 lb per gal. of water or 6% slurry at the maximum feed rate. Hydrated lime is usually used where maximum feed rates do not exceed 250 lb/hour. Hydrated lime and milk of lime will irritate the eyes, nose, and respiratory system and will dry the skin. Affected areas should be washed with water.

Storage. Information given for quicklime also applies to hydrated lime except that bin agitation must be provided. Bulk bin outlets should be provided with non-flooding rotary feeders.

Feed Equipment. Volumetric or gravimetric feeders may be used, but volumetric feeders are usually selected only for installations where comparatively low feed rates are required. Dilution does not appear to be important, therefore, control of the amount of water used in the feeding operation is not considered necessary. Inexpensive hydraulic jet agitation may be furnished in the wetting chamber of the feeder as an alternative to mechanical agitation. The jets should be sized for the available water supply pressure to obtain proper mixing.

Piping and Accessories. Piping and accessories as described for quicklime are also appropriate for hydrated lime.

Pacing and Controls. Controls as listed for dry alum apply to hydrated lime. Hydraulic jets should operate continuously and only shut off when the feeder is taken out of service. Control of the feed rate with pH as well as pacing with the plant flow may be used with hydrated lime as well as quicklime.

10.3.3 Estimated Initial Costs of Adding Lime Feed Facilities to Existing Wastewater Treatment Plants

Cost estimates of chemical storage and feeding equipment have been prepared based on the use of hydrated lime for a 1 mgd plant, and pebble quicklime for 10 and 100 mgd plants. Lime feed rates are based on a dosage of 150 mg/l and allow for peak rates of twice this capacity. Storage was provided for at least 15 days at the average rate although it may be desirable to provide 30 days for dry chemicals. This storage time is arbitrary and will vary at each installation depending on the distance to and the reliability of the source of chemical supply. Piping and buildings to house the feeding equipment are not included in the estimates. The estimated costs of steel bins with dust collector vents and filling accessories are included for all three plants. The chemical feeding equipment can be installed in existing buildings at many locations. In all estimates, the costs include an allowance for the contractor's installation, overhead, and profit, and an allowance of 20% of the construction cost for engineering and contingencies. The estimated costs for each size plant are:

Plant Size (mgd)	Estimated Base Bid ($)	Additional Cost of Flow and pH Controls ($)	Total Cost ($)
1.	15,800	3,500	19,300
10.	80,100	13,200	93,300
100.	332,200	92,700	424,900

Equipment for the 1.0 mgd plant includes two volumetric feeder systems with manually loaded bins, dissolving chambers and accessories. Equipment for the 10.0 mgd plant includes two gravimetric feeder-slaker systems. Steel bins are estimated to hold 1-1/2 truck loads of quicklime each, and it is assumed that the truck will be equipped with a pneumatic blower. Equipment for the 100.0 mgd plant includes four gravimetric feeder-slaker systems. The base estimate includes the bin gate and feeder-slaker accessories and controls.

Costs for the flow and pH controls include holding tanks with mixers, and slurry feeders with controls. Flow and pH metering instruments are not included.

10.4 Other Compounds for pH Adjustment

10.4.1 Soda Ash

Properties and Availability. Soda ash, Na_2CO_3, is available in two forms. Light soda ash has a bulk density range of 35 to 50 lb/ft^3 and a working density of 41 lb/ft^3. Dense soda ash has a density range of 60 to 76 lb/ft^3 and a working density of 63 lb/ft^3. The pH of a 1% solution of soda ash is 11.2.

The molecular weight of soda ash is 106. Commercial purity ranges from 98 to 99% Na_2CO_3. The viscosities of sodium carbonate solutions are given in Figure 10-8. Soda ash by itself is not particularly corrosive, but in the presence of lime and water, caustic soda is formed which is quite corrosive.

Soda ash is available in bulk by truck, box car and hopper car, and in 100 lb bags from the following locations:

Location	Manufacturer
CALIFORNIA	
Bartlett	PPG Industries, Inc.
Trona	American Potash and Chemical Corp.
Westend	Stauffer Chemical Co.
GEORGIA	
Brunswick	Allied Chemical Co.
LOUISIANA	
Baton Rouge	Allied Chemical Co.
Lake Charles	Olin Chemicals

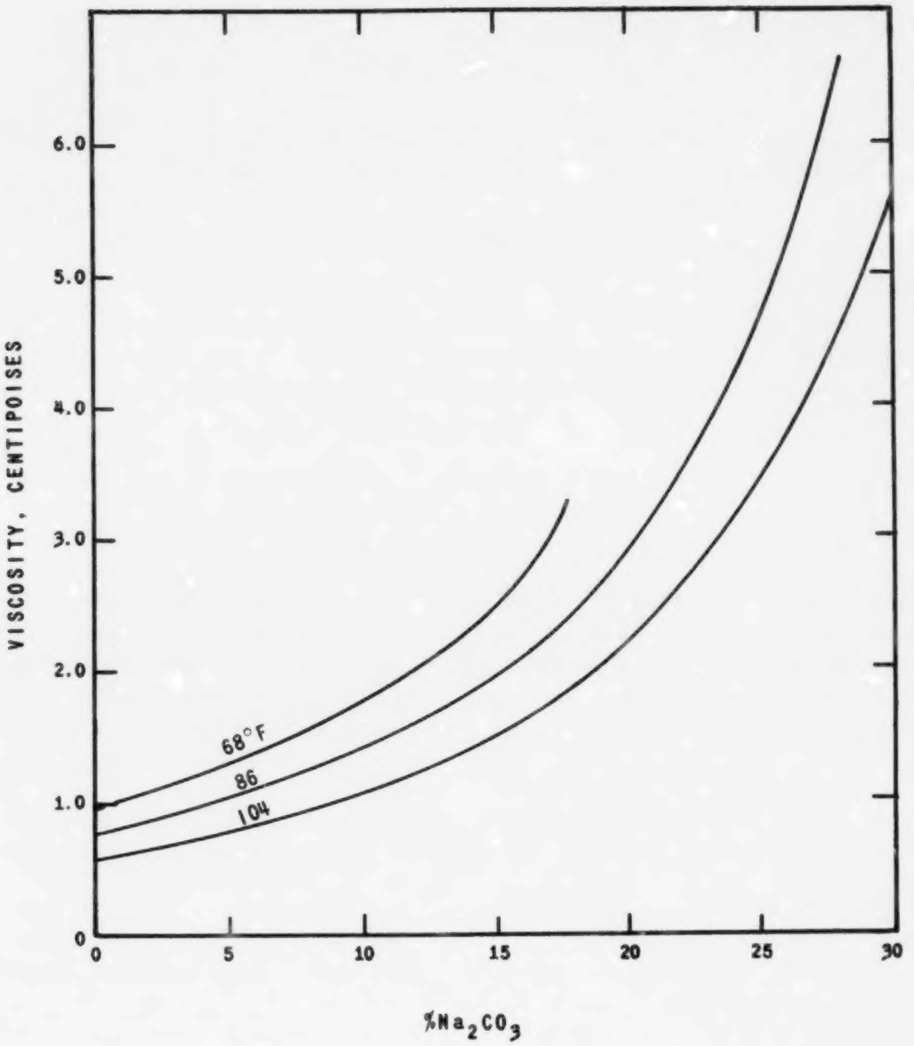


FIGURE 10-8 VISCOSITY OF SODA ASH SOLUTIONS
(Courtesy PPG Industries Inc., Chemical Div.)

Location	Manufacturer
MICHIGAN	
Wyandotte	Wyandotte Chemicals Corp.
NEW YORK	
Solvay	Allied Chemical Co.
OHIO	
Barberton	PPG Industries, Inc.
Painesville	Diamond Shamrock Chemical Co.
TEXAS	
Corpus Christi	PPG Industries, Inc.
WEST VIRGINIA	
Moundsville	Allied Chemical Co.
WYOMING	
Green River (3 plants)	Allied Chemical Co, FMC Corp., and Stauffer Chemical Corp.

The current price for soda ash ranges from $40 to $45/ton, F.O.B. the point of manufacture, however, prices vary substantially between manufacturers and should be obtained from the closest manufacturers or local distributors.

General Design Considerations. Dense soda ash is generally used in municipal applications because of superior handling characteristics. It has little dust, good flow characteristics, and will not arch in the bin or flood the feeder. It is relatively hard to dissolve and ample dissolver capacity must be provided. Normal practice calls for 0.5 lb of dense soda ash per gal. of water or a 6% solution retained for 20 minutes in the dissolver.

The dust and solution are irritating to the eyes, nose, lungs and skin and therefore general precautions should be observed and the affected areas should be washed promptly with water.

Storage. Soda ash is usually stored in steel bins and where pneumatic filling equipment is used, bins should be provided with dust collectors. Bulk and bagged soda ash tend to absorb atmospheric CO_2 and water forming the less active sodium bicarbonate ($NaHCO_3$). Material recommended for unloading facilities is steel.

Feeding Equipment. Feed equipment as described for dry alum is suitable for soda ash. Dissolving of soda ash may be hastened by the use of warm dissolving water. Mechanical or hydraulic jet mixing should be provided in the dissolver.

Piping and Accessories. Materials of construction for piping and accessories should be iron, steel, rubber, and plastics.

Pacing and Control. Controls as discussed for dry alum apply also to soda ash equipment.

10.4.2 Liquid Caustic Soda

Anhydrous caustic soda (NaOH) is available but its use is generally not considered practical in water and wastewater treatment applications. Consequently, only liquid caustic soda is discussed below.

Properties and Availability. Liquid caustic soda is shipped at two concentrations, 50% and 73% NaOH. The densities of the solutions as shipped are 12.76 lb/gal. for the 50% solution and 14.18 lb/gal. for the 73% solution. These solutions contain 6.38 lb/gal. NaOH and 10.34 lb/gal. NaOH, respectively. The crystallization temperature is 53° F for the 50% solution and 165° F for the 73% solution. The molecular weight of NaOH is 40. Viscosities of various caustic soda solutions are presented in Figure 10-9. The pH of a 1% solution of caustic soda is 12.9.

Truck load lots of 1,000 to 4,000 gal. are available in the 50% concentration only. Both shipping concentrations can be obtained in 8,000, 10,000, and 16,000 gal. car load lots. Tank cars can be unloaded through the dome eduction pipe using air pressure or through the bottom valve by gravity or by using air pressure or a pump. Trucks are usually unloaded by gravity or with air pressure or a truck mounted pump.

Major producers of caustic soda and their respective plant locations are listed in Table 10-3. The current price for liquid caustic soda ranges from about $65 to $70/ton (NaOH), F.O.B. the point of manufacture.

Table 10-3

LOCATION OF CAUSTIC SODA MANUFACTURING PLANTS

Location	Manufacturer
ALABAMA	
Lemoyne (Mobile)	Stauffer
McIntosh	Olin
Muscle Shoals	Diamond Shamrock
CALIFORNIA	
Pittsburg	Dow
DELAWARE	
Delaware City	Diamond Shamrock
GEORGIA	
Augusta	Olin
Brunswick	Allied
KANSAS	
Wichita	Vulcan
KENTUCKY	
Calvert City (2 plants)	Pennwalt, Goodrich

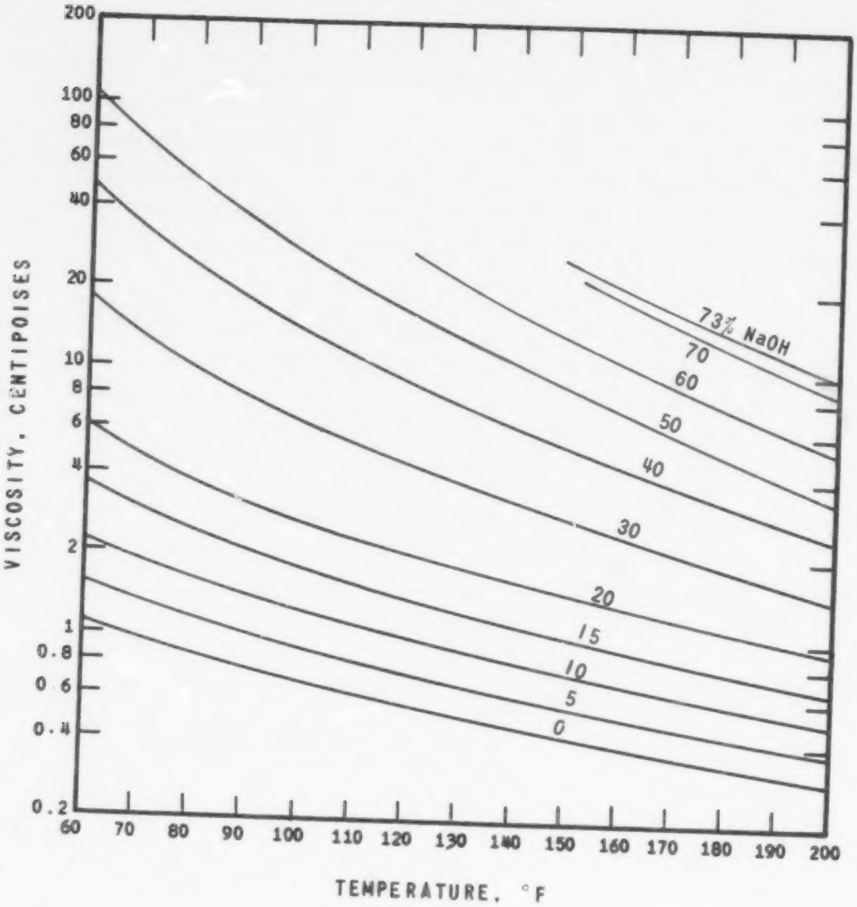


FIGURE 10-9 VISCOSITY OF CAUSTIC SODA SOLUTIONS
(Courtesy of Hooker Chemical Co.)

Table 10-3 (Continued)

Location	Manufacturer
LOUISIANA	
Baton Rouge	Allied
Geismar	Wyandotte
Lake Charles (2 plants)	PPG, Olin
Plaquemine	Dow
St. Gabriel	Stauffer
Taft	Hooker
MICHIGAN	
Midland	Dow
Montague	Hooker
Wyandotte (2 plants)	Pennwalt, Wyandotte
NEVADA	
Henderson	Stauffer
NEW JERSEY	
Linden	GAF
NEW YORK	
Niagara Falls (3 plants)	Hooker, Olin, Stauffer
Solvay	Allied
NORTH CAROLINA	
Acme	Allied
OHIO	
Barberton	PPG
Cleveland	Harshaw
Painesville	Diamond Shamrock
OREGON	
Portland	Pennwalt
TENNESSEE	
Charleston	Olin
TEXAS	
Corpus Christi	PPG
Deer Park (Houston)	Diamond Shamrock
Freeport	Dow
Port Neches	Jefferson
VIRGINIA	
Saltville	Olin
WASHINGTON	
Tacoma (2 plants)	Hooker, Pennwalt

Table 10-3 (Continued)

Location	Manufacturer
WEST VIRGINIA	
Moundsville	Allied
Natrium	PPG
South Charleston	FMC

Manufacturers and addresses

Allied Chemical
Solvay Process Division
40 Rector Street
New York, New York

Diamond Shamrock Chemical Co.
300 Union Commerce Building
Cleveland, Ohio 44115

Dow Chemical Co.
Abbott Road
Midland, Michigan 48640

GAF Corp. Chemical Division
140 West 51st Street
New York, New York

FMC Corporation
Inorganic Chemicals Div.
633 Third Avenue
New York, New York 10017

B. F. Goodrich Chemical Co.
3135 Euclid Avenue
Cleveland, Ohio

Harshaw Chemical Co.
1945 East 97th Street
Cleveland, Ohio 44106

Hooker Chemical Corp.
P. O. Box 344
Niagara Falls, New York 14302

Jefferson Chemical Co., Inc.
3336 Richmond Avenue
Houston, Texas 77006

Olin Corporation
Chemicals Division
745 Fifth Avenue
New York, New York

Pennwalt Corporation
Pennwalt Building
Three Penn Center
Philadelphia, Pa. 19102

PPG Industries, Inc.
Chemical Division
1 Gateway Center
Pittsburgh, Pa. 15222

Stauffer Chemical Co.
Industrial Chemical Div.
299 Park Avenue
New York, New York

Vulcan Materials Co.
Chemicals Division
P. O. Box 545-T
Wichita, Kansas 67201

Wyandotte Chemicals Corp.
Michigan Alkali Division
Wyandotte, Michigan

General Design Considerations. Liquid caustic soda is received in bulk shipments, transferred to storage and diluted as necessary for feeding to the points of application. Caustic soda is poisonous and is dangerous to handle. U. S. Department of Transportation Regulations for "White Label" materials must be observed. However, if handled properly caustic soda poses no particular industrial hazard. To avoid accidental spills, all pumps, valves, and lines should be checked regularly for leaks. Workmen should be thoroughly instructed in the precautions related to the handling of caustic soda. The eyes should be protected by goggles at all times when exposure to mist or splashing is possibie. Other parts of the body should be protected as necessary to prevent alkali burns. Areas exposed to caustic soda should be washed with copius amounts of water for 15 minutes to 2 hours. A physician should be called when exposure is severe. Caustic soda taken internally should be diluted with water or milk and then neutralized with dilute vinegar or fruit juice. Vomiting may occur spontaneously but should **not** be induced except on the advice of a physician.

Storage. Liquid caustic soda may be stored at the 50% concentration. However, at this solution strength, it crystallizes at 53°F. Therefore, storage tanks must be located indoors or provided with heating and suitable insulation if outdoors. Because of its relatively high crystallization temperature, liquid caustic soda is often diluted to a concentration of about 20% NaOH for storage. A 20% solution of NaOH has a crystallization temperature of about -20°F. Recommendations for dilution of both 73% and 50% solutions should be obtained from the manufacturer, because special considerations are necessary.

Storage tanks for liquid caustic soda should be provided with an air vent for gravity flow. The storage capacity should be equal to 1-1/2 times the largest expected delivery, with an allowance for dilution water, if used, or 2 weeks supply at the anticipated feed rate, whichever is greater. Tanks for storing 50% solution at a temperature between 75°F and 140°F may be constructed of mild steel. Storage temperatures above 140°F require more elaborate materials selection and are not recommended. Caustic soda will tend to pick up iron when stored in steel vessels for extended periods. Subject to temperature and solution strength limitations, rubber, 316 stainless steel, nickel, nickel alloys, or plastics may be used when iron contamination must be avoided.

Feeding Equipment. Further dilution of liquid caustic soda below the storage strength may be desirable for feeding by volumetric feeders. Feeding systems as described for liquid alum generally apply to caustic soda with appropriate selection of materials of construction. A typical system schematic is shown in Figure 10-10. Feeders will usually include materials such as ductile iron, stainless steels, rubber, and plastics.

Piping and Accessories. Transfer lines from the shipping unit to the storage tank should be spiral-wire-bound neoprene or rubber hose, solid steel pipe with swivel joints, or steel hose. Because caustic soda attacks glass, use of glass materials should be avoided. Other miscellaneous materials for use with liquid caustic soda feeding and handling equipment are listed below (7):

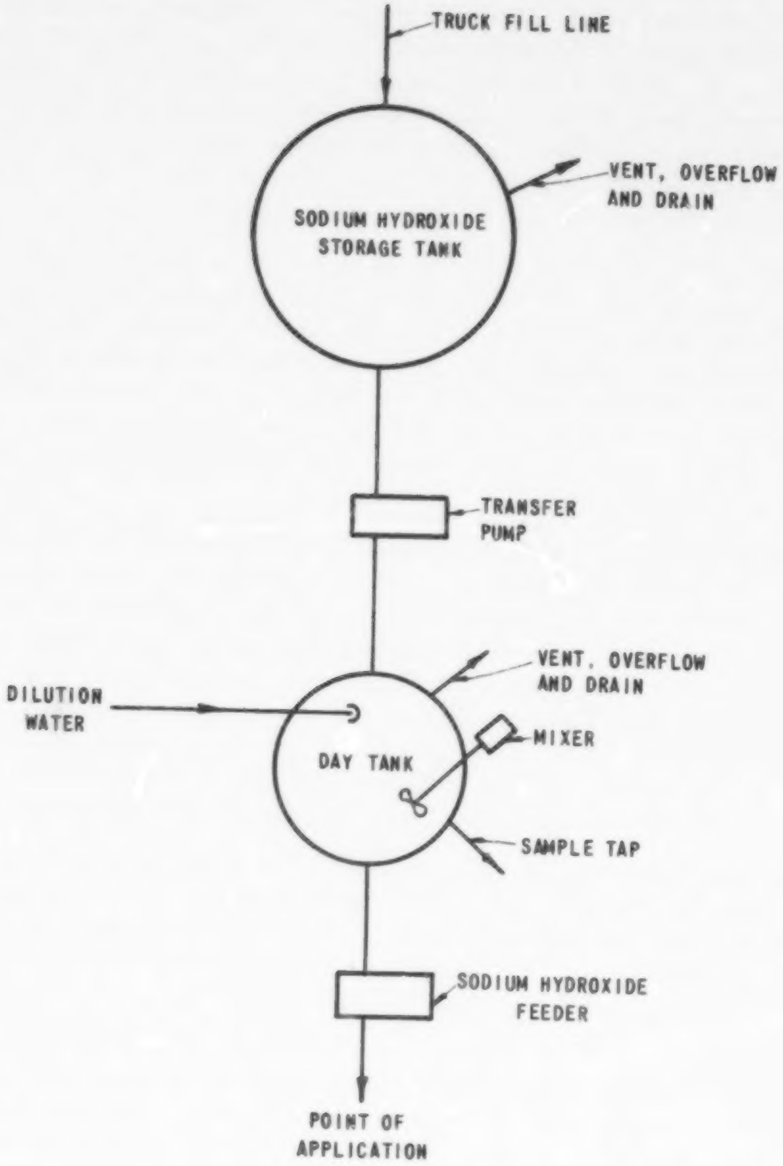

TRUCK FILL LINE
VENT, OVERFLOW AND DRAIN
SODIUM HYDROXIDE STORAGE TANK
TRANSFER PUMP
VENT, OVERFLOW AND DRAIN
DILUTION WATER
MIXER
DAY TANK
SAMPLE TAP
SODIUM HYDROXIDE FEEDER
POINT OF APPLICATION

FIGURE 10-10 TYPICAL CAUSTIC SODA FEED SYSTEM

Components	Recommended Materials for Use With 50% NaOH Up to 140° F
Rigid Pipe	Standard Weight Black Iron
Flexible Connections	Rigid Pipe with Ells or Swing Joints, Stainless Steel or Rubber Hose
Diluting Tees	Type 304 Stainless Steel
Fittings	Steel
Permanent Joints	Welded or Screwed Fittings
Unions	Screwed Steel
Valves – Non-leaking (Plug)	
Body	Steel
Plug	Type 304 Stainless Steel
Pumps (Centrifugal)	
Body	Steel
Impeller	Ni-Resist
Packing	Blue Asbestos
Storage Tanks	Steel

Pacing and Control. Controls as listed for liquid alum also apply to liquid caustic soda equipment.

10.5 Carbon Dioxide

Properties and Availability. Carbon dioxide, CO_2, is available for use in wastewater treatment plants in gas and liquid form. The molecular weight of CO_2 is 44. Dry CO_2 is not chemically active at normal temperatures and is a non-toxic safe chemical; however, the gas displaces oxygen and adequate ventilation of closed areas should be provided. Solutions of CO_2 in water are very reactive chemically and form carbonic acid. Saturated solutions of CO_2 have a pH of 4.0 at 68° F.

The gas form may be produced on the treatment plant site by scrubbing and compressing the combustion product of lime recalcining furnaces, sludge furnaces, or generators used principally for the production of CO_2 gas only. These generators are usually fired with combustible gases, fuel oil, or coke and have CO_2 yields as shown in Table 10-4 (8).

Table 10-4

CO_2 YIELDS OF COMMON FUELS

Fuel	Quantity	CO_2 Yield (lb)
Natural Gas	1,000 ft^3	115
Coke	1 lb	3
Kerosene	1 gal.	20
Fuel Oil (No. 2)	1 gal.	23
Propane	1,000 ft^3	141
Butane	1,000 ft^3	142

The gas forms, as generated at the plant site, usually have a CO_2 content of between 6% and 18% depending on the source and efficiency of the producing system.

The liquid form is available from commercial suppliers in 20 to 50 lb cylinders, 10 to 20 ton trucks and 30 to 50 ton rail cars. The commercial liquid form has a 99.5% CO_2 content.

Current prices range from $30/ton for 3,000 tons per year and over, to $68/ton for a quantity of 150 tons/year. These prices include an allowance for freight within a 100 mile radius of the point of manufacture. Another $6/ton may be added for each additional 100 miles to the point of destination. Major producers of commercial CO_2 are listed in Table 10-5.

Table 10-5

LOCATION OF CARBON DIOXIDE MANUFACTURING PLANTS

Location	Manufacturer
CALIFORNIA	
Watson (Los Angeles)	Liquid Carbonic
Oakland	Liquid Carbonic
Brea	Airco
Lathrop	Airco
Ventura	Cardox
Taft	Standard Oil
GEORGIA	
Augusta	Liquid Carbonic
ILLINOIS	
Morris	Cardox
Chicago	Airco
INDIANA	
Jeffersonville	Cardox
IOWA	
Clinton	Airco
Ft. Madison	Liquid Carbonic
Ft. Dodge	Liquid Carbonic
Muscatine	Publicker
KANSAS	
Dodge City	Liquid Carbonic
Lawrence	Airco (1972)
Lawrence	Cardox
KENTUCKY	
Doerun (Brandenburg)	Olin

Table 10-5 (Continued)

Location	Manufacturer
LOUISIANA	
New Orleans	Liquid Carbonic
Luling	Airco
MASSACHUSETTS	
Tewksbury	Liquid Carbonic
MISSISSIPPI	
Yazoo City	Airco
MISSOURI	
Kansas City	Airco
Le May (St. Louis)	Cardox
NEW JERSEY	
Paulsboro	Olin
Belleville	Liquid Carbonic
Deepwater	Airco
NEW MEXICO	
Bueyeros	SEC
Solana	SEC
Mosquero	SEC
NEW YORK	
Olean	Airco
OHIO	
Toledo	Cardox
Oregon (Toledo)	Liquid Carbonic
Lima	Airco
Huron	Cardox
PENNSYLVANIA	
Philadelphia	Liquid Carbonic
Thermice (Philadelphia)	Publicker
TENNESSEE	
Woodstock (Memphis)	Cardox
TEXAS	
Texas City	Liquid Carbonic
Dallas	Cardox
Dumas	Diamond Shamrock
VIRGINIA	
Hopewell	Airco
Saltville	Olin

Table 10-5 (Continued)

Location	Manufacturer
WASHINGTON	
Finley	Airco

Manufacturers and Addresses

Cardox Div. of Chemetron Corp.
Dept. TR
840 N. Michigan Avenue
Chicago, Illinois 60611

Airco Industrial Gases Div. of
Air Reduction Co.
575 Mountain Avenue
Murray Hill, N. J.

Liquid Carbonic Corporation
Dept. TR
135 S. LaSalle
Chicago, Illinois 60603

Olin Corporation
Chemicals Division
745 Fifth Avenue
New York, New York

Publicker Ind., Inc.
Walnut & Thomas
Philadelphia, Pa.

SEC
1033 Humble Place
El Paso, Texas 79987

Diamond Shamrock Chemical Co.
300 Union Commerce Building
Cleveland, Ohio 44115

Standard Oil Company of California
225 Bush
San Francisco, California 94104

General Design Considerations. Recovery of CO_2 from recalcining furnaces or incinerators is the least expensive source, but maintenance of stack gas systems is likely to be extensive because of the corrosive nature of the wet gas and the presence of particulate matter. Scrubber systems are required to clean the stack gas and especially designed gas compressors are necessary to provide the process injection pressure.

Pressure generators and submerged burners require less maintenance because the system pressure is established by compressors or blowers handling dry air or gas. On-site generating units have a limited range of CO_2 production as compared with the liquid storage and feed system, and therefore may require multiple units.

The liquid CO_2 storage and feed system generally includes a temperature-pressure controlled, bulk storage tank, an evaporation unit, and a gas feeder to meter the gas. Solution feeders, similar in construction to chlorinators, may be used to feed CO_2.

Storage. This section applies only to use of commercial liquid CO_2. Liquid system capacities encountered in wastewater treatment usually require on-site bulk storage units. Standard pre-packaged units are available, ranging in size from 3/8 to 50 tons capacity, and are furnished with temperature-pressure controls to maintain approximately 300 lb/in.^{2} at 0^{o}F conditions. The typical package unit contains refrigeration, vaporization, safety and control equipment. The units are well insulated and protected for outdoor location. The gas from the evaporation unit usually passes through two stages of pressure reduction before entering the gas feeder to prevent the formation of dry ice.

Feeding Equipment. Feeding systems for the stack gas source of CO_2 consist of simple valving arrangements, for admitting varying quantities of make-up air to the suction side of the constant volume compressors, or for venting excess gas on the compressor discharge. A typical system is described by Culp and Culp (9).

Pressure generators and submerged burners are regulated by valving arrangements on the fuel and air supply. Generation of CO_2 by combustion is usually difficult to control, requires frequent operator attention and demands considerable maintenance over the life of the equipment.

Commercial liquid carbon dioxide is becoming more generally used because of its high purity, the simplicity and range of feeding equipment, ease of control, and smaller, less expensive piping systems. After vaporization, the CO_2 with suitable metering and pressure reduction may be fed directly to the point of application as a gas. However, vacuum operated, solution type gas feeders are often preferred. Such feeders generally include safety devices and operating controls in a compact panel housing, with materials of construction suitable for CO_2 service. Absorption of CO_2 in the injector water supply approaches 100% when a ratio of 1.0 lb of gas to 60 gal. of water is maintained.

Piping and Accessories. Mild steel piping and accessories are suitable for use with cool, dry, carbon dioxide. Hot, moist gases, however, require the use of type 316 stainless steel or plastic materials. Plastics or FRP pipe are generally used for solution piping and diffusers. Diffusers should be submerged at least 8 ft, and preferably deeper, to assure complete absorption of the gas.

Pacing and Control. Standard instrument signals and control components can be used to pace or control carbon dioxide feed systems.

Using stack gas as the source of CO_2, the feed rate can be controlled by proper selection and operation of compressors, by manual control of vent or bleed valves, or by automatic control of such valves by a pH meter-controller system.

In commercial CO_2 feed systems, solution feeders may function as controllers and can be paced by instrument signals from pH monitors and plant flow meters.

In feeding commercial CO_2 directly to the point of application as a gas, a differential pressure transmitter and a control valve may function as the primary elements of a control system. Standard instrument signals may be used to pace or control the rate of CO_2 feed.

CO_2 generators are difficult to pace or control other than by manual or automatic operation of vent or bleed valves that waste a portion of the produced gas according to the plant requirements.

10.6 Polymers

10.6.1 Dry Polymers

Properties and Availability. Types of polymers vary widely in characteristics. Manufacturers should be consulted for properties, availability, and cost of the polymer being considered. References are available that indicate the types and characteristics of polymers available (10, 11). Bulk shipments are generally not desirable. Polymers are available in a variety of container or package sizes.

General Design Considerations. Dry polymer and water must be blended and mixed to obtain a recommended solution for efficient action. Solution concentrations vary from fractions of a per cent up. Preparation of the stock solution involves wetting of the dry material and usually an aging period prior to application. Solutions can be very viscous, and close attention should be paid to piping size and length and pump selections. Metered solution is usually diluted just prior to injection to the process to obtain better dispersion at the point of application.

Storage. General practice for storage of bagged dry chemicals should be observed. The bags should be stored in a dry, cool, low humidity area and used in proper rotation, i.e., first in, first out.

Solutions are generally stored in type 316 stainless steel, FRP, or plastic lined tanks.

Feed Equipment. Two types of systems are frequently combined to feed polymers. The solution preparation system includes a manual or automatic blending system with the polymer dispensed by hand or by a dry feeder to a wetting jet and then to a mixing-aging tank at a controlled ratio. The aged polymer is transported to a holding tank where metering pumps or rotodip feeders dispense the polymer to the process. A schematic of such a system is shown by Figure 10-11. It is generally advisable to keep the holding or storage time of polymer solutions to a minimum, 1 to 3 days or less, to prevent deterioration of the product.

Piping and Accessories. Selection must be made after determination of the polymer, however, type 316 stainless steel or plastics are generally used.

Pacing and Controls. Controls as listed for liquid alum apply to the control of liquid dispersing feeders.

The solution preparation system may be an automatic batching system, as shown by the schematic on Figure 10-12, that fills the holding tank with aged polymer as required by level probes. Such a system is usually provided only at large plants. Unitized solution preparation units are available, but have a limited capacity.

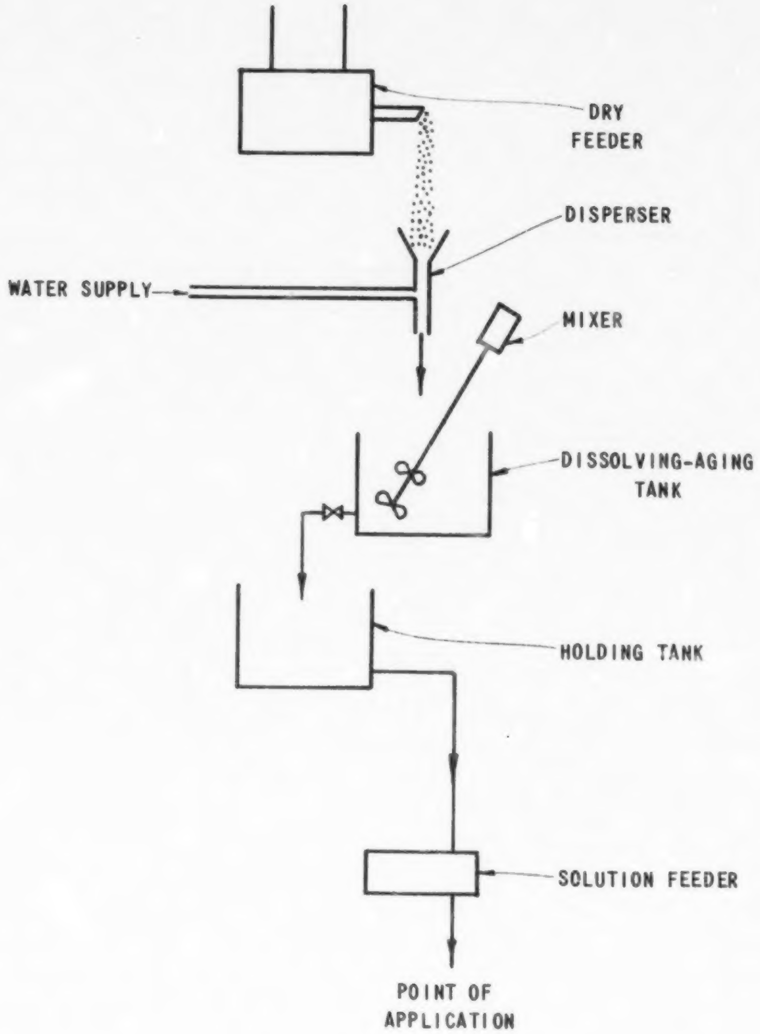
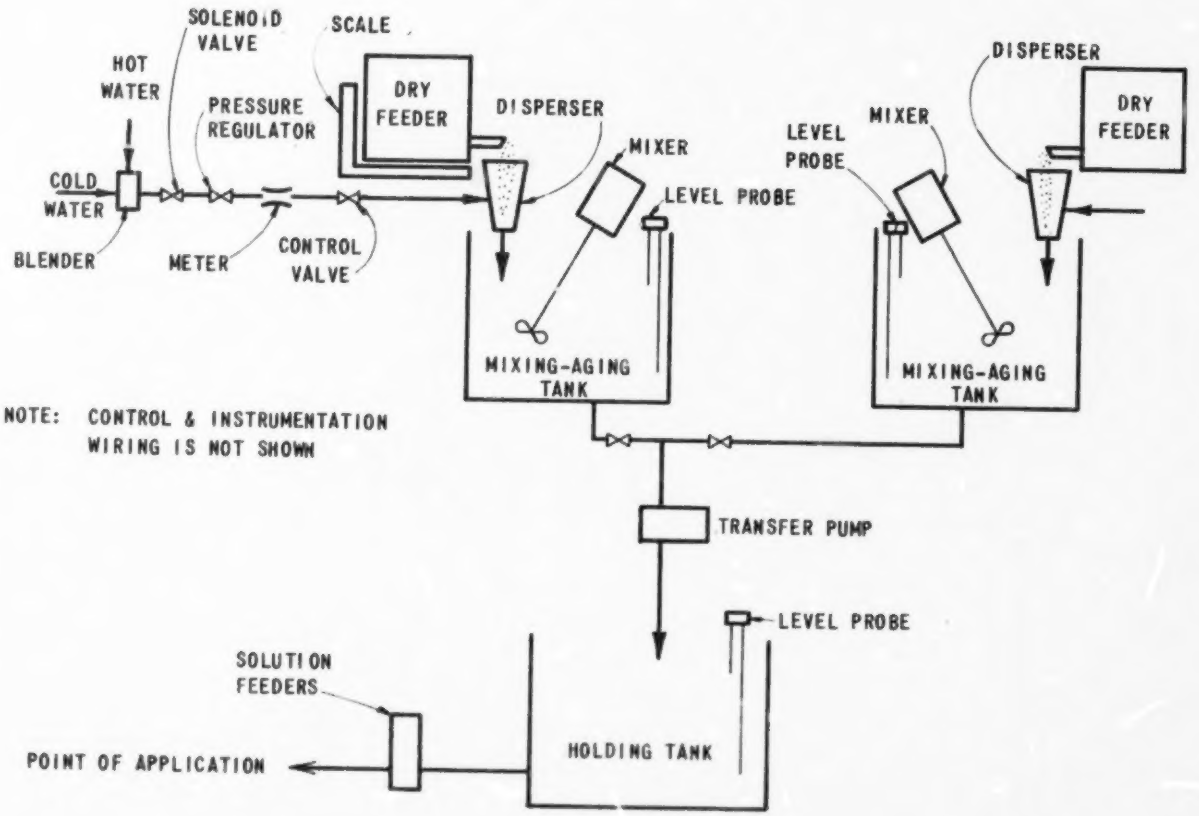

DRY FEEDER
DISPERSER
WATER SUPPLY
MIXER
DISSOLVING-AGING TANK
HOLDING TANK
SOLUTION FEEDER
POINT OF APPLICATION

FIGURE 10-11 TYPICAL SCHEMATIC OF A DRY POLYMER FEED SYSTEM


HOT WATER
SOLENOID VALVE
SCALE
DRY FEEDER
DISPERSER
MIXER
LEVEL PROBE
PRESSURE REGULATOR
COLD WATER
BLENDER
METER
CONTROL VALVE
MIXING-AGING TANK
LEVEL PROBE
MIXER
DISPERSER
DRY FEEDER
MIXING-AGING TANK
TRANSFER PUMP
LEVEL PROBE
SOLUTION FEEDERS
HOLDING TANK
POINT OF APPLICATION

NOTE: CONTROL & INSTRUMENTATION WIRING IS NOT SHOWN

FIGURE 10-12 TYPICAL AUTOMATIC POLYMER FEED SYSTEM FOR LARGE PLANTS

10.6.2 Liquid Polymers

Properties and Availability. As with dry polymers, there is a wide variety of products, and manufacturers should be consulted for specific information.

General Design Considerations. Liquid systems differ from the dry systems only in the equipment to blend the polymer with water to prepare the solution. Liquid solution preparation is usually a hand batching operation with manual filling of a mixing-aging tank with water and polymer.

The balance of the process is generally the same as described for dry polymers.

10.6.3 Estimated Initial Costs of Adding Polymer Feed Facilities to Existing Wastewater Treatment Plants

Cost estimates of solution preparation and feeding equipment have been prepared, based on the use of dry polymer, for 1, 10, and 100 mgd plants. Chemical feed equipment was sized to have a capacity sufficient to prepare and feed a 0.25% stock solution at a dosage of 1 mg/l. Piping and buildings to house the feeding equipment and store the bags are not included. At many locations, the chemical feeding equipment can be installed in existing buildings. All estimated costs include an allowance for the contractor's installation, overhead, and profit plus an allowance of 20% of the construction cost for engineering and contingencies. The estimated costs for each size plant are:

Plant Size (mgd)	Estimated Cost ($)
1	5,810
10	26,740
100	203,000

The system for the 1 mgd plant was based on the polymer being manually fed to the mixing tank, with mixing and solution feed equipment arranged for manual control. Two independent systems of tanks and feeders are included. The system for the 10 mgd plant includes two volumetric dry feeders discharging to two mixing tanks, arranged for batch control. The mixing tanks discharge to a single holding tank from which two solution feeders take their supply for application to the process. The system for the 100 mgd plant includes four volumetric dry feeders complete with steel day hoppers, dust collectors, bin gates and flexible connectors. The system operation is automatic for the preparation and transfer of aged polymer solution from four mixing tanks to two holding tanks. Ten solution feeders meter the polymer to the treatment process.

10.7 Rapid Mixing

For iron or aluminum compounds to be effective in phosphorus removal they must be intimately mixed with the wastewater. Chemical dispersion usually occurs in some type of mechanically-agitated rapid mix basin. (Mixing may also be accomplished by air agitation or a hydraulic jump.) Design of these basins is based on detention time and the velocity gradient in the basin. Walker (12) formulates the velocity gradient, G, as follows:

$$G = \sqrt{\frac{W}{\mu}}$$

Where:

$$W = \frac{\text{Water horsepower x 550}}{\text{Tank Volume (cubic feet)}}$$

μ = absolute viscosity of the fluid at the temperature involved, lb-sec/ft^2

Culp and Culp (9) recommend a velocity gradient of 300 ft/sec/ft or greater, a detention time of 15 to 60 seconds, and a power input to the mechanical mixers of 0.25 to 1.0 hp/mgd.

Others indicate that higher values of G, 700 to 1,000 ft/sec/ft are necessary for efficient blending. To attain a velocity gradient of 800 ft/sec/ft^2 at a detention time of 30 seconds and 68^{o}F, the water horsepower per mgd can be calculated as follows:

Volume of the mixing basin per mgd = 46.4 ft^3

$$G^2 = \frac{(HP_w)(550)/46.4}{0.2098 \times 10^{-4}}$$

$$(800)^2 \times 0.2098 \times 10^{-4} = 550\ HP_w/46.4$$

$$HP_w = 1.13$$

The required motor (brake) horsepower is:

$$BHP = \frac{HP_w}{\text{Efficiency}}$$

Assuming an efficiency of 75%, brake HP = $HP_w/0.75$ = 1.13/0.75 = 1.5. Therefore, 1.5 horsepower per mgd would be required to provide the desired mixing. Using Culps' figures of 300 ft/sec/ft, a 30 sec detention time, and the same temperature; the required HP_w = 0.16/mgd. Based on the same overall efficiency (75%), the required brake horsepower would be 0.21/mgd. Full scale studies should be conducted on rapid mix basins using the higher velocity gradients (700 to 1,000 ft/sec/ft) to investigate any advantages. The cost of mixers rated at a horsepower capable of producing the high G values does not appear to be prohibitive.

Recent investigations (13) emphasize the hydrolysis effect on trivalent aluminum and iron salts by dilution. Solutions of these salts will not hydrolyze as long as the solution pH is 3.0 or less. Upon dilution to a higher pH, hydrolysis occurs; that is, the trivalent aluminum and iron form hydrous oxide gels. Following this, the ions age through a series of trivalent species by polymerization until the neutral hydroxide form is attained. Hydrolysis takes place in less than 1/10 second and the neutral trivalent hydroxide species is reached within two or three minutes. Phosphorus precipitation capability of the aluminum or iron trivalent ion decreases as hydrolysis and aging progress; therefore, it is imperative to instantaneously disperse the chemical in the wastewater. The coagulant should be added to the vortex under a mechanical mixer or to the point of greatest turbulence in other types of mixers. This will assist in achieving the most rapid dispersion possible.

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Chapter 11

CHECKLISTS FOR REVIEW OF PLANS AND SPECIFICATIONS

11.1 Introduction

The processes outlined in this design manual have been developed for application to projects where improved effluent quality is needed or required. In some cases, these processes have had limited use in full-scale design of wastewater treatment facilities. Design engineers and reviewing authorities may not be completely familiar with the parameters to be considered in the preparation of plans and specifications for a given project. The following plan and specification review check sheet has been prepared to serve as a guide to the engineer in the design of proposed facilities which use the process or processes outlined in this manual. It will also be used by the Environmental Protection Agency, and may be used by State and local authorities in their review of projects for approval.

As with the case of any check sheet, its purpose is to fully consider all possible parameters for an individual process. For a given project all or part of the check sheet may be applicable and should be used with this fact in mind.

11.2 Phosphorus Removal by Mineral Addition Before the Primary Settler

11.2.1 Rapid Mix

Influent flow, mgd
Number of units
Detention time, seconds
Velocity gradient – G, ft/sec/ft
Mixer horsepower, hp
Freeboard, ft

11.2.2 Flocculator

Flow and recirculation, mgd
Number of units
Detention time, minutes
Drive horsepower, hp
Velocity gradient – G, ft/sec/ft
Velocity gradient x time – Gt
Freeboard, ft

11.2.3 Primary Settler

Flow and recycle, mgd
Number of units
Detention time, hours
Surface overflow rate, gpd/ft^2

- Weir rate, gpd/linear ft
- Sidewater depth, ft
- Freeboard, ft
- Solids removal efficiency, %
- Clarifier mechanism loading, lb/linear ft
- Solids pumping capacity, gpm

11.2.4 Chemical Storage and Feed System

- Dosage rate, mineral : phosphorus (molar ratio)
- Dosage rate, mg/l
- Bulk storage
  - Capacity
  - Construction material
- Pumps, transfer
  - Type
  - Capacity, gpm
  - Rated head, ft
- Dilution and feed tanks
  - Capacity, gal.
  - Construction material
- Agitators
  - Agitator horsepower, hp
  - Construction material
- Feeder or feed pumps
  - Type
  - Capacity, gph or lb/hour
  - Number of operating units
  - Number of standby units
- Control (flow, influent P)

11.2.5 Sludge Handling

- Sludge lines – capacity and head loss, gpm and ft
- Sludge pumps – capacity and head, gpm and ft

11.2.6 Gravity Thickener

- Number of units
- Solids loading, lb/ft^2/day
- Overflow rate, gpd/ft^2
- Detention time, hours
- Sidewater depth, ft
- Freeboard, ft
- Feed concentration, % dry solids
- Thickened concentration, % dry solids
- Solids recovery, %

Thickening mechanism – type
Mechanism loading, lb/linear ft

11.2.7 Centrifuge

Number of units
Type
Feed concentration, % dry solids
Underflow sludge concentration, % dry solids
Solids recovery, %
Dry solids capacity, lb/day
Total sludge, lb/day
Centrate return

11.2.8 Vacuum Filter

Number of units
Type
Feed concentration, % dry solids
Filter cake, % dry solids
Solids recovery, %
Dry solids capacity, lb/hour/ft^2
Total dry solids, lb/day
Filtrate return

11.2.9 Incinerator

Number of units
Sludge moisture, %
Sludge water, lb/hour
Sludge dry solids, lb/hour
Sludge total, lb/hour
Sludge volatile matter, % dry solids
Sludge ash content, % dry solids
Sludge heating value, Btu/lb
Allowable combustibles in ash, %
Operating schedule, hours/week
Auxiliary fuel
Maximum incinerator temperature, °F
Air pollution control devices
Maximum stack gas temperature, °F

11.3 Phosphorus Removal by Lime Addition Before the Primary Settler

11.3.1 Rapid Mix

See Section 11.2.1

11.3.2 Flocculator

See Section 11.2.2

11.3.3 Primary Settler

See Section 11.2.3

11.3.4 Lime Storage and Feed System

- Dosage rate, mg/l
- Bulk storage
  - Capacity, tons
  - Construction material
  - Bin dimensions, ft
  - Dust collector
- Lime feeders
  - Number of feeders
  - Type
  - Maximum capacity, lb/hr
  - Accessories
- Lime slaker
  - Number of slakers
  - Type
  - Maximum capacity, lb/hr
  - Slaking time, minutes
- Lime slurry holding tank
  - Capacity, gal.
  - Agitator horsepower, hp
- Lime slurry pumps
  - Type
  - Capacity, gpm
  - Number of operating units
  - Number of standby units
- Control (pH, flow)

11.3.5 Sludge Handling

See Section 11.2.5

11.3.6 Gravity Thickener

See Section 11.2.6

11.3.7 Centrifuge

See Section 11.2.7

11.3.8 Vacuum Filter

See Section 11.2.8

11.4 Phosphorus Removal in Trickling Filters by Mineral Addition

11.4.1 Rapid Mix (if necessary)

See Section 11.2.1

11.4.2 Flocculator (if necessary)

See Section 11.2.2

11.4.3 Final Settler

See Section 11.2.3

11.4.4 Chemical Storage and Feed System

Point of Chemical Addition
(See Section 11.2.4 for remaining parameters)

11.4.5 Sludge Handling

See Section 11.2.5

11.4.6 Gravity Thickener

See Section 11.2.6

11.4.7 Centrifuge

See Section 11.2.7

11.4.8 Vacuum Filter

See Section 11.2.8

11.4.9 Incinerator

See Section 11.2.9

11.5 Phosphorus Removal in Activated Sludge Plants by Mineral Addition

11.5.1 Rapid Mix (if necessary)

See Section 11.2.1

11.5.2 Flocculator (if necessary)

See Section 11.2.2

11.5.3 Final Settler

See Section 11.2.3

11.5.4 Chemical Storage and Feed System

Point of Chemical Addition
(See Section 11.2.4 for remaining parameters)

11.5.5 Sludge Handling

See Section 11.2.5

11.5.6 Gravity Thickener

See Section 11.2.6

11.5.7 Centrifuge

See Section 11.2.7

11.5.8 Vacuum Filter

See Section 11.2.8

11.5.9 Incinerator

See Section 11.2.9

11.6 Phosphorus Removal by Lime Treatment of Secondary Effluent

11.6.1 Rapid Mix

See Section 11.2.1

11.6.2 Flocculator

See Section 11.2.2

11.6.3 Tertiary Settler

See Section 11.2.3

11.6.4 Recarbonator

- Number of stages
- Basins
  - Detention time, minutes
  - Sidewater depth, ft
  - Inside depth, ft
- CO_2 source
- CO_2 dosage, lb/10^6gal.
- CO_2 dispersion
  - Location in basins
  - Type
  - Control

        - Depth of submergence, ft
        - Size of diffuser orifices, in.

11.6.5 Multimedia Filter

- Average flow at steady state, gpm
- Filtration rate, gpm/ft^2
- Total surface area, ft^2
- Available head, ft
- Media
    - Type
    - Depth, in.
    - Effective size, mm
    - Uniformity coefficient
    - Support material
- Underdrains
    - Type
    - Head loss, ft
- Freeboard, ft
- Backwash
    - Maximum rate, gpm/ft^2
    - % Expansion
    - Length of cycle, minutes
    - Washwater troughs
        - Number
        - Distance from top of media to bottom of trough, ft
    - Surface wash
        - Rate, gpm/ft^2
        - Duration, minutes
    - Air wash
        - Rate ft^3/minute/ft^2
        - Air pressure, lb/in^2
        - Duration, minutes

11.6.6 Lime Storage and Feed System

See Section 11.3.4

11.6.7 Sludge Handling

See Section 11.2.5

11.6.8 Gravity Thickener

See Section 11.2.6

11.6.9 Centrifuge

See Section 11.2.7

11.6.10 Vacuum Filter

See Section 11.2.8

11.6.11 Recalcining Furnace

Number of units
Type
Operating schedule, hours/week
Furnace diameter, ft
Number of hearths
Composition of the feed sludge
    Volatile suspended solids, %
    Ash, %
    Fuel value of volatile suspended solids, Btu/lb
Sludge concentration, % dry solids
Dry solids, lb/day
Water, lb/day
Auxiliary fuel
Maximum recalcination temperature, °F
Air pollution control devices
Maximum stack gas temperature, °F
Recalcined lime cooler

11.6.12 Recalcined Lime Storage

Capacity, tons
Construction material
Bin dimensions, ft
Dust collector

11.7 Phosphorus Removal by Mineral Addition to Secondary Effluent

11.7.1 Rapid Mix

See Section 11.2.1

11.7.2 Flocculator

See Section 11.2.2

11.7.3 Tertiary Settler

See Section 11.2.3

11.7.4 Multimedia Filter

See Section 11.6.5

11.7.5 Chemical Storage and Feed System

See Section 11.2.4

11.7.6 Sludge Handling

See Section 11.2.5

11.7.7 Gravity Thickener

See Section 11.2.6

11.7.8 Centrifuge

See Section 11.2.7

11.7.9 Vacuum Filter

See Section 11.2.8

11.7.10 Incinerator

See Section 11.2.9

APPENDIX

Abbreviations

The following are abbreviations of terms and dimensions used in the manual:

Terms

BOD	Biochemical oxygen demand
BOD_5	Five day BOD
COD	Chemical oxygen demand
DO	Dissolved oxygen
FRP	Fiberglas reinforced plastic
MLSS	Mixed liquor suspended solids
SS	Suspended solids
VSS	Volatile suspended solids

Dimensions

gpd	Gallons per day
gph	Gallons per hour
gpm	Gallons per minute
hp	Horsepower
JTU	Jackson turbidity units
mgd	Million gallons per day
rpm	Revolutions per minute

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